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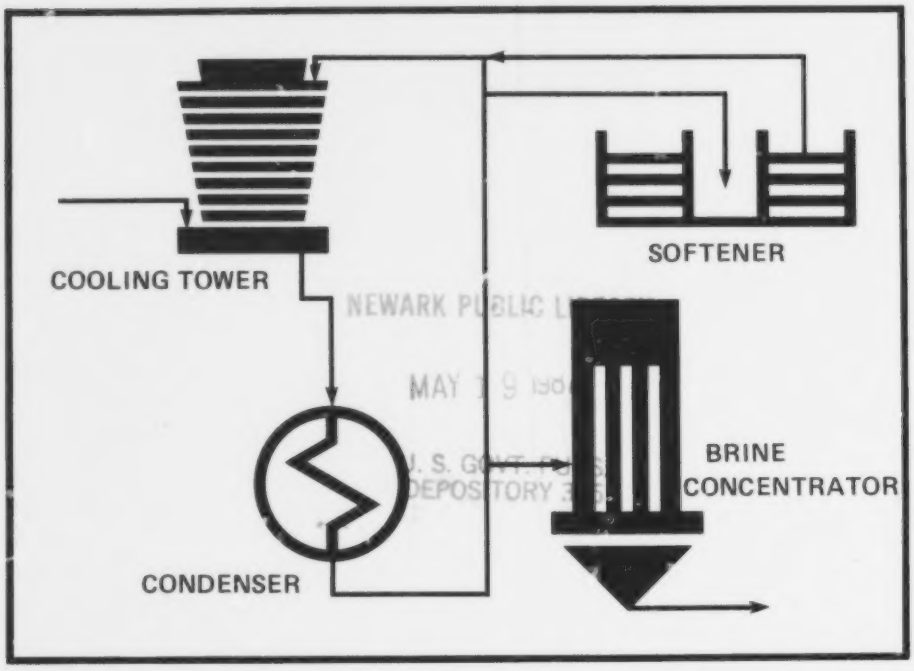
STUDY OF SALINE WATER USE AT THE ETIWANDA GENERATING STATION



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FINAL REPORT

APRIL 1987

Prepared for
U.S. DEPARTMENT OF THE INTERIOR
Bureau of Reclamation

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FINAL REPORT

TASK 3

BASIC AGREEMENT NUMBER 4-AB-81-03600

STUDY OF SALINE WATER USE AT THE ETIWANDA GENERATING STATION

PREPARED FOR

UNITED STATES DEPARTMENT OF THE INTERIOR
BUREAU OF RECLAMATION

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MARCH 16, 1987

7-2090 (4-81)
Water and Power

TECHNICAL REPORT STANDARD TITLE PAGE

1. REPORT NO.	2. GOVERNMENT ACCESSION NO.	3. RECIPIENT'S CATALOG NO.

4. TITLE AND SUBTITLE
Final Report - Study of Saline Water Use at the Etiwanda Generating Station

5. REPORT DATE
March 1987

6. PERFORMING ORGANIZATION CODE

7. AUTHOR(S)
Jack K. Laughlin

8. PERFORMING ORGANIZATION REPORT NO.

9. PERFORMING ORGANIZATION NAME AND ADDRESS

10. WORK UNIT NO.

11. CONTRACT OR GRANT NO.

13. TYPE OF REPORT AND PERIOD COVERED

12. SPONSORING AGENCY NAME AND ADDRESS
Bureau of Reclamation
Engineering and Research Center
Denver, CO 80225

14. SPONSORING AGENCY CODE

15. SUPPLEMENTARY NOTES
This report was prepared for the Bureau of Reclamation through an agreement with Jack K. Laughlin, Consulting Engineer.

16. ABSTRACT

This report presents the results of a study to evaluate water treatment process options for the use of Palo Verde Irrigation Outfall Drain as a major water supply for a hypothetical coal-fired powerplant located in the Mojave Desert between Blythe, California, and Las Vegas, Nevada. The study was done in conjunction with a field test of ion exchange softening technology conducted at the Etiwanda Generating Station near San Bernardino, California. Data from the field test program were used to support the design and cost evaluations of ion exchange softening technology in the study.

Sidestream lime/soda softening was the most cost-effective process option for treating Palo Verde Irrigation Outfall Drain water quality.

Ion exchange softening technology, using cooling tower blowdown for regeneration appears to be technically viable based on short-term testing, and is likely to be competitive in cost for waters high in hardness, high in sodium, and low in silica.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS-- saline water/cooling towers/salinity control/economic analysis/powerplant cooling/Etiwanda Generating Station/powerplant water supply/softening/process selection/

b. IDENTIFIERS--

c. COSATI Field/Group COWRR: SRIM

18. DISTRIBUTION STATEMENT	19. SECURITY CLASS (THIS REPORT)	21. NO. OF PAGES
Available from the National Technical Information Service, Operations Division, 5285 Port Royal Road, Springfield, Virginia 22161.	UNCLASSIFIED	86
	20. SECURITY CLASS (THIS PAGE) UNCLASSIFIED	22. PRICE

GPO 831-316

PREFACE

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EXECUTIVE SUMMARY

This report presents the results of Task 3 of Bureau of Reclamation Contract 4-AB-81-03600 which evaluated water treatment process options applicable to the use of Palo Verde Irrigation Outfall Drain as a water supply for a hypothetical coal-fired power plant located in the Mojave Desert between Blythe, California and Las Vegas, Nevada. The Task 3 study was performed in conjunction with a field test of ion exchange softening technology conducted at the Etiwanda Generating Station near San Bernardino, California. Data from the field test program were used to support the design and cost evaluations of ion exchange softening technology in the study.

The comparison of water treatment process options was based on a two-unit coal-fired powerplant with 600 megawatts of total net generating capacity. The reference powerplant assumed California regulatory criteria, and included the use of wet limestone scrubbers for sulfur dioxide control and zero discharge of liquid wastes. The Palo Verde Irrigation Outfall Drain water supply was chosen because of the availability of a complete water quality data base, an adequate quantity for a 600 megawatt powerplant and because the water quality was suitable for evaluation of ion exchange softening technology for cooling water makeup treatment.

Five water treatment process options, were developed for technical and economic comparisons. These options included:

I Ion exchange softening of cooling tower makeup, sidestream caustic/soda softening of the circulating water system, vapor compression evaporators for wastewater concentration and reuse, and evaporation ponds for final wastewater disposal

II Sidestream lime/soda softening of the circulating water system, vapor compression evaporators for wastewater concentration and reuse, and evaporation ponds for final wastewater disposal

III No makeup or sidestream treatment of cooling water, no evaporators for wastewater concentration and reuse, and evaporations ponds only for wastewater disposal

IV No makeup or sidestream treatment of cooling water, vapor compression evaporators for wastewater concentration and reuse, and evaporation ponds for final wastewater disposal

V Makeup lime/soda softening for cooling water treatment, vapor compression evaporators for wastewater concentration and reuse, and evaporation ponds for final wastewater disposal

Data from the Etiwanda Field Test Program were used to supplement the design of the ion exchange softening system. The Etiwanda field test results showed that ion exchange softening, using cooling tower blowdown for regeneration, was technically viable for both fluidized bed upflow regeneration and for rapid downflow regeneration. The process design for Option I used the more conservative upflow configuration.

Appraisal-level cost estimates, at an accuracy of plus or minus 25%, were developed for capital, chemical usage, operating and maintenance labor and materials and power cost. Levelized cost comparisons for the five process options were developed using microcomputer spreadsheet models and economic factors provided by Southern California Edison Company. The results of the economic analyses are summarized as follows:

Cost Comparison of Process Options
(Levelized Cost in $1000)

Option	I	II	III	IV	V
Capital	6247	5369	25369	7901	7998
Chemicals	3006	2397	701	561	2780
Labor & Repair	1436	1213	2162	1066	1403
Power	3153	2032	302	5202	3988
Total	13842	11011	28534	14730	16169

These comparative costs show that Option II (sidestream softening) is lowest in total levelized cost, followed by Option I (ion exchange softening), which is approximately 26% higher.

An economic sensitivity analysis was performed to examine the effect on varying capital cost, chemical cost, O&M labor and materials cost and power cost from 75% to 150% of the base case costs. The sensitivity analysis showed that only chemical cost changed the relative ranking of the options. Option I slips to third place if chemical costs reach 150% of the base case, illustrating the impact of high chemical usage for sidestream silica control.

The overall results of the Task 3 study produced the following conclusions:

- Sidestream lime/soda softening was the most cost-effective process option for treating Palo Verde Irrigation Outfall Drain water quality.

- Ion exchange softening technology, using cooling tower blowdown for regeneration, appears to be technically viable based on short-term testing, and is likely to be competitive in cost for waters which are high in hardness, high in sodium and low in silica.

- Wastewater evaporators are more cost-effective than evaporation ponds for disposal of recyclable wastewaters, and provide the additional benefit of a high quality water supply for full or partial makeup for other plant systems including condensate and flue gas desulfurization. Conceptual-level cost estimates for the use of waste heat steam as an alternative to electric-driven vapor compressors for wastewater evaporators showed significant cost savings potential, provided the issues of in-plant steam supply system design and availability are fully determined.

- Both ion exchange and lime/soda softening technologies allow operation of the powerplant circulating water system at high cycles of concentration, which can produce a highly corrosive environment for

cooling system materials

Based on these conclusions, the following recommendations are made:

- Ion exchange softening technology should be tested on a demonstration plant scale if site-specific power plant design criteria and water supply quality show an economic advantage. The test program would be enhanced by involvement of ion exchange resin manufacturers and ion exchange equipment suppliers.

- Sidestream lime/soda softening technologies should be demonstrated on cooling systems using high salinity makeup waters to develop accurate technical and economic data.

- Waste heat evaporators for concentration and reuse of powerplant wastewaters should be evaluated on a design-level basis in sufficient detail to develop accurate cost comparisons with vapor compression evaporator technology.

- Powerplant cooling water system materials of construction should be evaluated for application in high chloride and high total dissolved solids operating environments to account for plant operation on saline water sources at high cycles of cooling tower concentration.

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Section 1

INTRODUCTION

BACKGROUND

Under the Colorado River Water Quality Improvement Program (CRWQIP), the Bureau of Reclamation is studying opportunities for collecting and using saline water sources for beneficial industrial purposes [1] (references are shown in Section 10). This study evaluates water treatment process options applicable to the use of Palo Verde Irrigation Outfall Drain as a water supply for a hypothetical coal-fired power plant located in the Mojave Desert between Blythe, California and Las Vegas, Nevada. The Palo Verde Irrigation Outfall Drain is a tributary stream of the Colorado River system with an average salinity of approximately 1,800 mg/l.

The Task 3 study was performed in conjunction with a field test of ion exchange softening technology conducted at the Etiwanda Generating Station near San Bernardino, California. Field test program participants included the Bureau of Reclamation (Reclamation), Southern California Edison Company (SCE), Pacific Gas and Electric Company (PG&E), the California Department of Water Resources (DWR), the U.S. Environmental Projection Agency (EPA), and Sephton Water Technology (SWT).

The purpose of the field test program was to determine the technical viability of ion exchange softening for power plant cooling tower makeup water treatment, using cooling tower blowdown to regenerate the ion exchange softening system. The primary focus of the study was to compare the cost of ion exchange softening technology with alternative water treatment process options for the hypothetical power plant. Palo Verde Irrigation Outfall Drain water quality was simulated in the field test unit. Data from the field test program were applied to the final design and the economic analysis of the ion exchange softening option.

STUDY APPROACH

The study was organized as a team effort including Jack Laughlin, Rick Schroeder, Roger Jorden and Bob Hamilton of Stearns Catalytic Corporation. Technical support for design and evaluation of ion exchange softening equipment was provided by John Kaakinen of the Bureau of Reclamation. The study involved the following major steps.

- Development of a design water analysis for the Palo Verde Irrigation Outfall drain
- Development of a reference coal-fired power plant design based on sizing, siting and regulatory criteria provided by SCE and PGandE
- Review and selection of commercially viable saline water use technologies

- Development of water treatment process design options including ion exchange softening, and alternative technologies
- Economic modeling and levelized cost comparison of the selected water treatment process options
- Monitoring of the field test program at Etiwanda, and application of the field test data to process and economic updates of the ion exchange softening option
- Economic sensitivity analysis of major cost variables
- Comparison of the levelized in-plant cost-effectiveness for salt removal from the Colorado River system for each of the water treatment process options

The study included only differential water system costs within the plant boundaries. All study work was performed on an appraisal-level basis with an accuracy of plus or minus 25%.

Section 2

POWER PLANT DESIGN

A coal-fired power plant design was developed as a basis for comparing the Etiwanda Field Test Program water treatment process with alternative saline water treatment configurations. The reference power plant characteristics were selected in coordination with Southern California Edison (SCE) and Pacific Gas and Electric (PG&E).

PLANT DESCRIPTION

The reference power plant includes two generating units each with 330 megawatts of gross power capacity, producing a total net generating capacity of 600 megawatts. Each generating unit is a base-loaded sub-critical pulverized coal-fired type unit with all auxiliaries necessary for startup, operation and emission control.

It is assumed that the first unit will begin commercial operation in 1995, followed by the second unit in 1996.

Plant Siting

The plant siting assumes a desert environment similar to the California-Nevada-Arizona border area between Interstate Highways 15 and 10. This environment is similar in characteristics to the existing Mohave Generating Station site and the sites studied for the Harry Allen, Ivanpah and Sundesert Stations.

California regulations are assumed for air quality control, wastewater disposal and solid waste disposal. The plant is designed for zero liquid discharge.

Coal Quality

A typical coal quality was selected from PG&E data. Because the reference power plant will not use a specific coal source, an analysis of a sample of coal from the Sage Point area in eastern Carbon County, Utah was selected as being typical of coal delivered to the site. These coal quality data are shown in Appendix B-1. This coal quality has a heating value of 12,000 btu/lb, an ash content of 7%, a sulfur content of 0.5% and an as-received moisture content of 7%.

Raw Water Supply

Palo Verde Irrigation Outfall Drain water is assumed as the primary raw water supply for the plant. The Palo Verde Irrigation Outfall Drain is located along the Colorado River near Blythe, California and collects irrigation drainage from adjacent lands. This water source was studied by Stone and Webster during the period of 1975 to 1979 as a potential cooling water supply source for the Sundesert Nuclear Plant [2].

PLANT WATER BALANCE

The overall plant water balance varies with the individual process options. The major catagories of the plant water balance include evaporation losses, discharge flows and total makeup requirements.

Evaporation Losses

The average evaporation losses for the plant operating systems at 100% load are shown in Table 2-2. Cooling tower evaporation losses vary with load, but all other evaporation losses are assumed constant.

TABLE 2-2

PLANT EVAPORATION LOSSES
(gpm)

Cooling Towers	5,900
Ash Systems	20
FGD System	500
Total	6,420

System Discharge Flows

A summary of the principal system discharge flows for the plant are listed in Table 2-3. Variable flows are dependent on the individual water treatment process options.

Plant Makeup Water Requirements

Makeup water requirements are the sum of the evaporation losses, losses to solids disposal -- minus the water recovered through recycle.

The total plant makeup water requirement with no recycle ranges from 7,000 to 8,300 gpm average flow at 100% load, depending on the process option.

TABLE 2-3

AVERAGE SYSTEM DISCHARGE FLOWS
(gpm)

System	Flow	Treatment Category
Condensate System		
Filter Backwash	25	Recycle
Boiler Blowdown	25	Recycle
High Quality Drains	10	Recycle
Regeneration Waste	50-125	Desalting and Recycle, Evaporation Ponds, Solids Drying
Cooling Tower Blowdown	Variable	Front End Treatment, Sidestream Treatment, Desalting and Recycle, Evaporation Ponds, Solids Drying
Ash System	100	Direct Recycle, Desalting and Recycle, Evaporation Ponds, Solids Drying
Plant Drains	100	Oil Removal and Recycle Desalting and Recycle, Evaporation Ponds, Solids Drying
FGD System	Variable	Ponding, Landfill

Section 3

SELECTION OF WATER TREATMENT PROCESS OPTIONS

The objective of this activity was the selection of water/wastewater treatment process options for the technical and economic comparison with the ion exchange softening process tested at Etiwanda Station. This activity included a review of commercially available and emerging saline water use technologies, and the the selection of five process options for appraisal-level process design and economic analysis.

SALINE WATER USE TECHNOLOGY REVIEW

Applicable saline water technologies for raw water and wastewater treatment and final wastewater disposal were selected from the results of the Bureau of Reclamation Task 1 and Task 2 studies [3][4] and from power industry experiences. Emerging technologies with potential cost savings were included. The technologies reviewed in the Task 1 and 2 evaluations included:

- Ion exchange softening (Ix)
- Lime/soda softening
- Binary cooling towers
- Spiral reverse osmosis (RO)
- Seeded reverse osmosis (SRO)
- Electrodialysis (ED)
- Freeze concentration
- Vertical tube vapor compression evaporators (VCE)
- Vertical tube foaming evaporators (VTFE)
- Horizontal tube spray film evaporators
- Cooling tower blowdown evaporators
- Solar evaporation ponds
- Spray-enhanced solar evaporation ponds
- Salt gradient ponds

Based on the plant design features, the site location and the Palo Verde Irrigation Outfall Drain water quality, ion exchange softening, lime/soda softening, reverse osmosis, electrodialysis, vertical tube vapor compression evaporators, vertical tube foaming evaporators and solar evaporation ponds were selected for incorporation into the water treatment process designs.

Ion Exchange Softening

The application of ion exchange softening of high hardness water for power plant cooling tower make-up offers a potential cost savings through the use of cooling tower blowdown for regeneration. Ion exchange (zeolite) softening is a long established technique for removing hardness (calcium and magnesium) from water. Historically, applications have ranged from home water softeners to industrial boiler make-up and small municipal treatment plants. Unfortunately, the volume of salt (NaCl) required and the corresponding volume of regenerant waste produced have precluded the application of ion exchange softening treatment of large volume flows. Modifications to the conventional

sodium chloride regenerating cycle in the ion exchange process have been developed and tested by the Bureau of Reclamation and others. These modifications have focused on utilizing reject desalting brine or cooling tower blowdown to regenerate the resin bed. Since these streams already exist, their cost and waste volume impact on the plant should be less than for a conventional sodium chloride brine regenerated system. These waste streams provide the sodium for regeneration as sodium sulfate rather than the traditional sodium chloride. The waste streams also contain a high concentration of other ions along with significant levels of suspended solids. Tests to date have shown the other ions to have minimal effect on the regeneration efficiency. Problems with high suspended solids in the regenerant have been avoided by conducting the regeneration upflow rather than downflow. These streams are low in NaCl concentration (1% to 3%) as compared with a saturated brine solution (10%). However, the "free" nature of the regenerant should provide favorable economics.

Ion exchange softening removes hardness only. Silica, another major scale forming constituent in most waters, is not removed. The ideal application for this approach is on water supplies with high hardness, low silica, and low suspended solids. If, however, the silica content of the water becomes a limiting factor at the cooling towers, silica control in the circulating water is a necessary cost penalty.

In the 1970's the Bureau of Reclamation sponsored and directed field testing of ion exchange prior to reverse osmosis and electrodialysis desalting at sites near Yuma, Arizona and LaVerkin Springs, Utah [5][6]. The feedwater at the Yuma site had a TDS level near 3,000 mg/l. The test unit was successfully regenerated with all waste brine concentrations tested, which ranged from 20,000 to 50,000 mg/l TDS. The LaVerkin Springs test unit achieved performance equaling that obtainable with conventional sodium chloride regeneration. This is attributed to the essentially unlimited volume of brine available and regeneration of the resin through multiple recycling of the regenerant.

Ion exchange softening of power plant saline water was tested in the Firebaugh Project sponsored by the California Department of Water Resources (DWR), Southern California Edison Company (SCE), Pacific Gas and Electric Company (PG&E) and the Los Angeles Department of Water Power (LADWP), and the Electric Power Research Institute (EPRI) [7]. San Joaquin Valley, California, irrigation water was the test water. The process test sequence included ion exchange softening of cooling tower make-up, cooling tower operation at 20 cycles of concentration, sidestream silica control with alum dosing, and regeneration of the ion exchange softener resin with the 20-fold concentrate.

The test results indicated technical viability, but the limited scope of the Firebaugh test program did not confirm performance and economic viability. Additional test was recommended.

The application of ion exchange softening for treatment of irrigation drainage water was tested by the California Department of Water Resources at the Los Banos test facility in Central California. In this test, ion exchange softening was used as pretreatment for a reverse osmosis desalting system. A goal of the Los Banos test was to produce drinking quality water from the San Luis

Drain which is primarily irrigation water with a salinity of approximately 9,000 mg/l. The water was first clarified, filtered, and ion exchange softened prior to entering the RO unit. Reject from the RO unit was used as regenerant to the ion exchange system. The reject was a 10-fold concentration over the feed water which resulted in a regenerant strength of approximately 10%. The ion exchange system had a nominal capacity of 400,000 GPD and consisted of four 20-square foot columns. Each column had a 5-foot deep resin bed and was regenerated upflow at 50% to 70% bed expansion. Downflow regeneration was not tested. Although there were some initial equipment problems, the system was able to sustain several thousand hours of operation. No gypsum precipitation was noted in the resin beds. The presence of scale inhibitors in the RO reject may have helped. There was however, some gypsum scale noted in the storage tanks downstream. This may have provided an indication of the time-dependent nature of gypsum scale formation. The system was able to achieve good hardness reduction closely paralleling published performance data. The resin showed no unusual attrition over the test period. Work at the Los Banos site was completed in December, 1986. However, additional test work will commence at Mendota in the fall of 1987.

By coincidence, the feedwater to the Los Banos test unit closely parallels the LaVerkin Springs feedwater. For this reason it is expected ion exchange performance at Los Banos should approximate that at LaVerkin Springs. The similarity of the feedwaters prompted CH2MHill to use performance data from LaVerkin Springs to project ion exchange performance as reverse osmosis pretreatment for a study of methods to treat San Luis Drain water [8]. This study indicated a favorable cost advantage of ion exchange softening over lime/soda softening as pretreatment.

The Etiwanda field test program evaluated the ion exchange softening process under actual plant operating conditions using simulated Palo Verde Irrigation Outfall Drain water quality. The field test unit included an ion exchange softening system operated in conjunction with a sidestream caustic/soda softener for silica control. The ion exchange softening system was regenerated with cooling tower blowdown using two regeneration configurations: (1) fluidized bed upflow and (2) rapid fixed bed downflow. Both configurations operated successfully. The Etiwanda field test data were used to augment the design and cost evaluation for process option using ion exchange softening in the Task 3 study.

Lime/Soda Softening

Lime softening is one of the most common make-up treatment techniques. It involves removal of impurities by chemical reaction with combinations of lime, soda ash, and caustic with the following effects:

- Calcium hardness precipitates as calcium carbonate and magnesium hardness as magnesium hydroxide; calcium precipitates orthophosphate as calcium phosphate
- Bicarbonate alkalinity is first converted to carbonate form and then removed as calcium carbonate

- Silica comes out as an insoluble silicon-magnesium complex when it reacts with, or is adsorbed on, solid magnesium hydroxide

- Turbidity from suspended solids and magnesium hydroxide and calcium carbonate sludge are removed by settling and/or filtration

A more recent application of lime softening is for cooling tower sidestream treatment. The sidestream is taken from the recirculating line downstream from the condenser to take advantage of the increased efficiency of silica removal by magnesium hydroxide and the lower calcium carbonate solubility at the higher temperature prevailing at this point in the system. Generally a sidestream softener with associated chemical feed equipment is several times smaller than a make-up softener for the same generating unit. This is accomplished by the concentrating effect of the circulating water system which provides the softener unit with a higher concentration of dissolved solids than would be encountered in the make-up position.

Sidestream lime/soda softening has emerged through recent power plant experiences at Martin Drake Station in Colorado, Coolwater Station in California and North Valmy Station in Nevada as a viable application. Additional experiences in domestic industrial plants and in foreign power plants further supports the viability of sidestream softening.

Combined make-up and sidestream lime/soda softening has shown promise in a number of power plant water management studies and in field tests. EPRI pilot plant tests conducted at Comanche Station in Colorado and at North Valmy Station in Nevada confirmed the technical potential of the process [9][10]. Tests conducted by Stone and Webster for San Diego Gas and Electric Company using the Palo Verde Irrigation Outfall Drain water also confirmed the technical and economic potential of the process [2]. The combined make-up and sidestream lime/soda softening application remains to be demonstrated in a full scale power plant system.

Southwest Public Service Company (SWPSCO) conducted pilot tests for using secondary municipal sewage as cooling tower make-up where the calcium phosphate solubility product limited the cycles of concentration [11]. They considered make-up softening, sidestream softening, and combined make-up and sidestream softening. They also evaluated lime, caustic, and lime/caustic mixes. They found the optimum choice to be combined make-up and sidestream softening in separate vessels. Lime was used in the make-up unit with a lime/caustic mixture in the sidestream unit. Plans are to incorporate this system into the final plant design.

Lime/soda softening for raw water make-up treatment, sidestream treatment and combined make-up and sidestream treatment was selected for process screening in the study based on demonstrated performance and economic potential.

Reverse Osmosis

Reverse osmosis is a treatment process for the production of high purity water by passing water through a semipermeable membrane permitting the passage of water but not the dissolved solids. The driving force is provided by a high

pressure pumping allowing the membranes reject both cations and anions and unionized species such as silica.

Reverse osmosis (RO) is applicable to raw water treatment, sidestream treatment of the cooling tower circulating water and to the concentration of wastewaters. Present RO technology includes spiral wound elements, hollow fine fiber elements and tubular elements. Each is applicable to the treatment of raw water. Spiral wound and tubular elements are generally considered more applicable to wastewater applications because of their higher resistance to fouling.

For either raw water source, RO would qualify as a commercially proven technology for boiler make-up water treatment, with numerous power plant references. For circulating water make-up treatment, the volume of waste produced by RO precludes its application.

Sidestream RO treatment's first major application is at the Bayswater Station in Australia. The Bayswater Station is presently in startup. Based on operating experiences on power plant wastewaters, the sidestream application of RO should be technically viable, but has not yet been commercially demonstrated.

The use of spiral wound RO as a wastewater concentration process has been demonstrated at the R.D. Nixon Station in Colorado, the San Juan Station in New Mexico, and the Martin Lake and Monticello Stations in Texas. The combined experiences of these power plants have shown that RO is a commercially viable wastewater treatment process, provided proper design, operating and maintenance approaches are used. Spiral RO was selected for evaluation in the Task 3 study because of its demonstrated performance for wastewater concentration.

Seeded RO, using a calcium sulfate seed slurry process to prevent scaling, is a new technology with potential make-up water and wastewater applications. Seeded RO requires minimal pretreatment compared to conventional RO in applications where calcium and silica reductions are required in the feed water. Seeded RO has been demonstrated through laboratory tests in the U.S., a pilot test at the Hunter Generating Station in Utah and through pilot tests conducted in South Africa on mine wastewaters.

The laboratory and pilot tests showed that the seeded RO process is technically viable, but that membrane life is sensitive to specific wastewater quality. Seeded RO was not selected for application in the water treatment process designs because it has not been demonstrated on a commercial scale and accurate capital and oporating cost data is not yet available.

Electrodialysis

Electrodialysis (ED) is similar to reverse osmosis in that semipermeable membranes are used for removal of dissolved solids. The primary differences lie in the characteristics of the membranes and the driving force. An ED system contains a stack of alternately placed cation and anion permeable membranes with spacers which allow flow parallel to the membranes. A DC cur-

rent provides the motive force for the selective ion migration through these membranes. Modifications to the basic ED process include Electrodialysis Reversal (EDR) where the polarity of the DC current is periodically reversed automatically which provides for on-stream cleaning of the system and a reduction in the fouling potential and lower chemical requirements.

Electrodialysis (ED or EDR) is applicable to raw water treatment, sidestream treatment of the cooling tower circulating water and wastewater concentration. For the reference raw water chemistry, ED is a commercially proven process for make-up water treatment. ED has been installed at approximately five domestic power plants for demineralizer pretreatment, and at a larger number of foreign plants for similar use.

To our knowledge, electrodialysis as a sidestream treatment process for power plant cooling water systems has been studied, but not tested. The technical viability for this application would be similar to applying ED in a wastewater concentration mode.

An ED wastewater concentration commercial application is installed at the ESCOM power plant complex in South Africa. In this application, ED is used as a pre-concentrator for a wastewater evaporator. No operating data was available from this installation.

ED has not been used domestically for power plant wastewater concentration. Analogous domestic applications include the concentration of cooling tower blowdown at an oil refinery, the concentration of saline water at the Bureau of Reclamation La Verkin Springs Test Facility, and the concentration of RO blowdown at the Los Banos test facility. These experiences have shown that ED for wastewater concentration has technical and economic promise.

Vapor Compression Evaporators

Vapor compression evaporation (VCE) concentration of power plant wastewaters is commercially proven technology. These units are long tube, vertical, falling film, evaporators using a vapor compression cycle. The wastewater is fed to the evaporator, mixed with a concentrated slurry, and continually recirculated to the top of the evaporator where it is distributed to the inside wall of each tube as a thin film. As the thin film falls down the inside of the tubes, water is converted to steam. The steam is compressed by an electrically driven compressor to provide the energy to evaporate the water from the brine. The condensed steam becomes a high quality water for recycled use in the plant.

Over two dozen VCE units, ranging in feedwater capacities from 150 to 600 gallons per minute, have been installed at power stations and have demonstrated reliable performance using a seeded slurry process for calcium sulfate scale control. In addition to concentrating power plant wastewaters, the VCE systems have produced a high quality distillate suitable for condensate make-up with minimal treatment.

VCE's were selected for the development of process options because of proven performance and applicability.

Vertical Tube Foaming Evaporators

Vertical tube foaming evaporators differ from VCE's in that a surfactant, or foaming agent, is used to enhance heat transfer performance and also to serve as a dispersant during seeded slurry operation. VTFE evaporators can be designed to operate with either vapor compressors, turbine extraction steam or waste heat steam as an energy supply. VTFE technology has been under development and demonstration since the mid-1960's. VTFE was demonstrated at the Firebaugh test program in the mid-1970's. An operational test of a 40,000 gallon per day VTFE driven by waste heat steam was subsequently sponsored by Southern California Edison Company and EPA at the Etiwanda Generating Station. This test program confirmed the technical viability of the foaming seeded slurry process for cooling tower blowdown concentration using low temperature turbine exhaust steam. Like the VCE, the VTFE produces distillate which can be used for power plant condensate make-up with minimal treatment.

Based on the demonstrated performance of VTFE technology for concentration of cooling tower blowdown, and engineering reports indicating potential cost savings for waste heat VTFE application, VTFE technology was selected for evaluation in the Task 3 study.

Evaporation Ponds

The use of solar evaporation ponds for final disposal of power plant wastewaters is commercially proven technology in areas where positive annual average net evaporation rates are present. For the study, evaporation ponds are applicable to all wastewater disposal options.

SCREENING OF PROCESS OPTIONS

From the review of saline water use technologies, seven process configurations were developed for screening. The process screening was based on an integrated power plant design approach incorporating the technologies which were technically viable and had the best potential for cost-effective use of saline water for power plant make-up. The screening studies eliminated membrane processes for cooling tower raw water treatment or for wastewater concentration, but retained the technologies of ion exchange softening, lime or caustic/soda softening, VTFE, VCE and evaporation ponds. The results of process screening are summarized as follows:

Ix/VTFE/VTFE/Evaporation Ponds

The process originally proposed for the Etiwanda field test program included ion exchange softening of cooling tower make-up, cooling tower circulating water silica control by sidestream treatment with alum, concentration of cooling tower blowdown by a steam-driven VTFE, and regeneration of the ion exchange softener by VTFE blowdown. In a full zero discharge plant design, additional wastewater concentration and disposal capabilities would be required.

With the Palo Verde Irrigation Outfall design water quality, the use of both sidestream silica control and a VTFE for regeneration of the ion exchange sof-

tener would not be economical. In the Ix/VTFE/VTFE/Evaporation Pond option, an evaluation was made of using the VTFE for final concentration of ion exchange regenerant rather than the cooling tower. In this option, cooling tower circulating water concentration is controlled to a silica limit of 150 mg/l (7 cycles). Chloride levels are maintained within the limits of normal circulating water system materials. Cooling tower blowdown, at a rate of approximately 850 gpm, is fed to the VTFE system where it is further concentrated to approximately 30 cycles (as referenced to the raw water supply.) The VTFE blowdown is then used to regenerate the ion exchange softener. Other plant wastewaters are concentrated in another steam-driven VTFE system, and non-recyclable wastewaters and sludges are discharged to evaporation ponds. High quality water from the VTFE systems is used for condensate make-up.

This process configuration has the advantage of operating the cooling water system within established chloride limits, but has the major disadvantage of the high capital cost associated with using the VTFE as a dedicated regeneration system for the Ix softener. Preliminary economic analysis indicated that the levelized cost of this option was not competitive.

Ix/Sidestream Silica Treatment/VTFE/Evaporation Ponds

This process option represents the actual configuration used in the Etiwanda field test loop and serves as the reference process design for economic comparisons. In this configuration, the ion exchange softeners are regenerated by cooling tower blowdown rather than by the VTFE blowdown. Sidestream silica treatment is used in the cooling tower circulating water system enabling the cooling towers to operate at the required 30 cycles of concentration to regenerate the Ix softeners, while keeping the silica levels in the circulating water loop below 150 mg/l. Other plant wastewaters are concentrated in a steam-driven VTFE system. Non-recyclable wastewaters and sludges are discharged to evaporation ponds. High quality water from the VTFE units is used as condensate make-up.

Alternative sidestream silica treatment processes were considered for this option, including sidestream caustic soda softening and sidestream treatment using chemical additives such as alum (as used at Firebaugh), magnesium oxide, and other standard silica reduction chemicals. Sidestream caustic soda softening provides silica control with the added benefit of calcium reduction for the circulating water. If partial calcium reduction in the sidestream loop is used, only partial ion exchange softening of the cooling tower raw water makeup is required, reducing the size and capital cost of the ion exchange softener.

Laboratory experiments conducted by Brown and Caldwell for the Etiwanda field test program indicated that sidestream caustic soda treatment was preferable for the simulated Palo Verde Irrigation Outfall water quality. Sidestream caustic soda softening was subsequently assumed in the process designs for the ion exchange softening options.

VTFE technology was selected for wastewater concentration in this option because a VTFE unit was to be included in the Etiwanda field test unit hardware.

Lime/Soda Sidestream Softening/VCE/Evaporation Ponds

In this configuration, raw water is used for cooling tower make-up and sidestream lime/soda softening of the cooling tower circulating water is used to minimize cooling tower blowdown. Vapor compression evaporators are used for concentration of recyclable wastewaters, and evaporation ponds are used for disposal of non-recyclable wastewaters and sludges. High quality water from the wastewater evaporators is used as condensate make-up.

VCE evaporators are assumed in place of VTFE evaporators because the economic factors for VCE technology have been well-established through numerous power plant installations.

The use of sidestream softening permits the control of calcium, magnesium, silica, phosphate and suspended solids in the circulating water loop. With the reference raw water quality, high chloride levels are produced in the circulating water loop including cooling towers, main condenser, bearing cooling water heat exchangers and hydrogen coolers. The technical and economic viability of this process configuration was demonstrated in the Task 1 and Task 2 studies for Hunter and Jim Bridger Stations. Other than the question of high chloride impacts on the circulating water loop, this process can be considered commercially proven and potentially very cost-effective. For these reasons, the sidestream lime/soda softening option was selected for economic analysis.

Evaporation Ponds

In this configuration, no make-up or sidestream treatment of the cooling water loop is used. Cooling tower cycles are controlled to the scaling limits of the raw water chemistry and within chloride limits applicable to standard circulating water system materials of construction. All non-recyclable wastewaters and sludges are discharged to evaporation ponds.

This option is the least complex of any of the configurations. It has the disadvantages of a large area of evaporation ponds and no recycled water available. Because this option represents established power plant technology and was studied by SCE as a comparison with VTFE technology, it was selected for economic analysis.

VCE/Evaporation Ponds

In this configuration, no raw water treatment or sidestream treatment of the cooling water loop is used. Recyclable wastewaters are concentrated in a vapor compression evaporator. High quality water from the evaporator is used as make-up to the condensate, circulating water, ash and FGD systems.

The use of VCE combined with evaporation ponds is a standard method for achieving zero discharge in many western power plants. This combination, low in operating complexity and generally more cost-effective than evaporation ponds alone, was selected for economic analysis.

Lime/Soda Raw Water Softening/VCE/Evaporation Ponds

In this configuration, raw water is treated by lime/soda softening for use as cooling tower make-up. Recyclable wastewaters are concentrated in vapor compression evaporators. Non-recyclable wastewaters and sludges are discharged to evaporation ponds. High quality water from the wastewater evaporators is used as make-up for the condensate, ash and FGD systems.

This configuration is well proven in power plants. In the Task 1 study for Hunter Station, lime/soda raw water softening was found to be cost-competitive with the best process options.

Combined Make-Up and Sidestream Lime/Soda Softening

In this configuration, both make-up and sidestream lime/soda softening are used. Make-Up and sidestream softening can be combined in one vessel or carried out in separate vessels. Recyclable wastewaters are concentrated in vapor compression evaporators. Non-recyclable wastewaters and sludges are discharged to evaporation ponds. High quality water from the wastewater evaporators is used as condensate make-up.

The combined make-up/sidestream softening process provides partial make-up softening by using high pH effluent from the sidestream softener. Chemical costs can be potentially reduced, but capital costs are greater than sidestream softening alone. As with sidestream softening, the concentrations of chlorides in the circulating loop adversely impacts materials costs.

As mentioned in the screening of technologies, the combined make-up/sidestream lime/soda softening process has been tested in two pilot plant programs and found to be technically feasible. It has also been evaluated in a number of power plant water management studies and found to be cost-competitive with make-up water treatment or sidestream treatment applied separately. The studies and fields tests conducted by Stone and Webster on the Palo Verde Irrigation Outfall Drain water for the Sundesert Nuclear Plant concluded that combined make-up/sidestream softening was the most cost-effective process.

Although the combined make-up/sidestream softening process has economic potential, it was not selected for economic analysis because of the lack of demonstrated commercial operation and because of the limited number of options which could be included in the economic analysis.

Application of RO and ED for Pre-Concentration of Wastewater

In each of the concepts involving wastewater evaporators, both spiral RO, and ED can be applied as pre-concentrators of the feed to the wastewater evaporators. If minimal pretreatment were required, the membrane processes would have a substantially lower capital and operating cost than either steam-driven or vapor compression evaporators.

In the reference power plant, cooling tower blowdown will be concentrated to the scaling limits of calcium sulfate and silica. This means that substantial pretreatment would be required to prevent scaling in the spiral RO or ED sys-

tems. Preliminary process and economic analysis showed that membrane processes were not economically competitive for wastewater concentration.

Summary of Process Options for Economic Evaluation

The following five process configurations were selected for economic analysis:

1. Ix/Sidestream Silica Control/VTFE/Evaporation Ponds

2. Sidestream Softening/VCE/Evaporation Ponds

3. Evaporation Ponds

4. VCE/Evaporation Ponds

5. Raw Water Softening/VCE/Evaporation Ponds

Section 4

DESIGN OF WATER TREATMENT PROCESS OPTIONS

Appraisal level process designs were developed for the five selected options based on the reference power plant design. Process flow diagrams and material balances were developed for each option and are at the end of this section.

The raw water supply system is assumed to be the same for each option. The raw water supply system includes polymer injection of the Palo Verde Irrigation Outfall water upstream of the Raw Water Reservoir, settling of solids in the reservoir and delivery of raw water to the power plant at 90 psig via a pump station located at the reservoir.

The quantity of raw water required for each option is adjusted for the amount of water recovered for recycle in the station. The recovered water from Options I, II, IV and V satisfies a portion or 100% of the condensate, FGD and ash system make-up requirements of the reference power plant. Option IV does not recover wastewater suitable for make-up to these systems.

FGD system scrubber waste flow is assumed to be the same for each option and is blended with the other wastewater streams for treatment. The disposal of final concentrated wastewaters utilizes lined evaporation ponds for all options. The evaporation pond design assumes a minimum depth of 15 feet for liquid and solids containment, and a single high density polyethylene liner of 80 mils bottom thickness and 60 mils side thickness. Net wastewater evaporation rate for the site is 2 gpm per acre, after adjustment for salinity effect. The evaporation ponds include underflow collection and monitoring systems as required to meet regulatory criteria.

Option I - Ix Softening/Sidestream Silica/VTFE/Evaporation Ponds

Option I includes cooling tower make-up softened by ion exchange, regeneration of the ion exchange softeners with concentrated cooling tower blowdown, concentration of other recyclable plant wastewaters by VTFE's, and final wastewater disposal in evaporation ponds. The process uses sidestream silica treatment in the circulating water system to minimize cooling tower blowdown and produce the circulating water concentration necessary for regenerating the ion exchange softener. The process flow diagram and material balance for Option I are presented in Figure 4-1 and Table 4-1.

The cooling tower make-up is distributed to 12 ion exchange vessels operating in parallel and having a total feed capacity of 10,000 gpm. The average flow requirement is 7,200 gpm. The excess capacity is necessary to support the regeneration schedule and redundancy required to maintain 90% availability. Each ion exchange vessel is 10 feet in diameter and 15 feet high. Approximately 4,500 cubic feet of resin is needed to fill the system.

The ion exchange softener effluent is collected in a 2.5 million gallon collection basin from which the cooling tower make-up water is pumped to the cooling tower basins. The water is treated at the basins with sulfuric acid

to control calcium carbonate scaling. Inhibitor for calcium sulfate scale control and chlorine for microbiological control in the circulating water system are also added.

The cooling towers are operated at 30 cycles of concentration, resulting in a total dissolved solids concentration of 52,000 mg/l and a suspended solids loading of 300 mg/l. Silica control below 150 mg/l in the circulating water is accomplished with two 600 gpm contact clarifiers in the sidestream position, operating in parallel. The average total feed flow to the clarifiers is estimated at 850 gpm. Caustic soda and soda ash are used for magnesium precipitation, silica absorption, and partial calcium softening. The clarifier feed is collected in a 1 million gallon surge pond. The effluent is transferred to the collection basin after the ion exchange system for blending with the cooling tower make-up. The sludge is transferred to the evaporation ponds for final disposal. Circulating water is blown down to an agitated, 150,000 gpm regenerant tank that supplies the ion exchange system at an average rate of 180 gpm. Regeneration is upflow with a fluidized bed resin.

The regenerant waste from the ion exchange softening system is expected to be supersaturated in calcium sulfate. Two clarifiers, operating in parallel, are required to desupersaturate the calcium sulfate in the spent regenerant solution. The effluent is transferred to the regenerant tanks for reuse. Calcium sulfate sludge is discharged to the evaporation ponds.

The recyclable plant wastewaters including the ash system discharge, miscellaneous drains, FGD wastewater, ion exchange waste effluent and condensate make-up system regeneration waste are collected in a 5 million gallon wastewater surge pond. Two 350 gpm waste heat VTFE units operating in parallel are used to concentrate the wastewaters. The average processing rate is 535 gpm, from which 484 gpm of distillate is reclaimed for reuse in the station. At the average processing rate, total steam requirements are 242,000 lbs per hour and the cooling requirements for distillate condensation is 34,000 gpm. The VTFE system design assumes a 25% contingency factor for VTFE availability and system upsets, but does not account for interruptions in steam supply such as unit outage.

The remaining 51 gpm of brine is discharged, along with the 20 gpm of ion exchange waste and 11 gpm of silica sludge, to 41 acres of evaporation ponds.

Option II - Sidestream Softening/VCE/Evaporation Ponds

Option II includes sidestream lime/soda softening of the circulating water, VCE for concentration of recyclable wastewaters and evaporation ponds for disposal of sludges and non-recyclable wastewaters. The process flow diagram and material balance for Option II is presented in Figure 4-2 and Table 4-2. Cooling tower make-up, at an average flow of 6,100 gpm, is delivered directly from the reservoir to the cooling tower basins. Sulfuric acid, scale inhibitor and chlorine are injected directly to the cooling tower basins for control of scale and microbiological activity.

The sidestream softening system includes a 1 million gallon softener surge pond that collects the average cooling tower blowdown flow of 650 gpm and

feeds two 1000 gpm softener/clarifiers operating in parallel. A 1 million gallon neutralization pond collects the softener effluent for proportional distribution back to the cooling towers. Storage and delivery systems are provided for lime, soda ash and settling polymers. The cooling towers are operated at 28 cycles of concentration based on a silica limit of 150 mg/l and a calcium limit of 600 mg/l. The circulating water TDS is 44,000 mg/l and is high in chlorides at approximately 11,000 mg/l. The circulating water suspended solids, at 300 mg/l, exceed the normal limit by 150 mg/l and may require special design consideration in low velocity heat exchangers. Sidestream softener effluent is controlled at 100 mg/l calcium and 75 mg/l silica through the addition of 85% lime and 98% soda ash. Circulating water alkalinity is controlled by sulfuric acid injection.

The plant wastewaters are collected in a 5 million gallon surge pond for processing by VCE's. The total inflow of 555 gpm is fed to two 350 gpm VCE units operating in parallel. The VCE product rate is 495 gpm with 60 gpm of waste brine produced. The VCE and sidestream softener sludge wastes, totaling 81 gpm, are discharged to 41 acres of evaporation ponds.

The recovered water satifies the condensate make-up and ash system needs but falls short of the FGD system make-up needs. Raw water used for FGD system make-up is treated with sulfuric acid to control calcium carbonate scale.

Option III - Evaporation Ponds

Option III is the least complicated process and provides no make-up and sidestream treatment, no VCE'S, and uses only evaporation ponds for disposal of final sludges and wastewaters. The process flow diagram and material balance for Option III are shown on Figure 4-3 and Table 4-3.

Raw water treatment for the cooling water, ash and FGD systems is minimal, and includes sulfuric acid addition to control calcium carbonate scale, scale inhibitor for calcium sulfate scale and chlorine for microbiological control.

All wastewaters are discharged to evaporation ponds for final disposal. Based on an inflow of 1,693 gpm, 845 acres of ponds are needed to maintain zero discharge for the life of the plant.

Option IV - VCE/Evaporation Ponds

Option IV includes minimal water treatment, treatment of recyclable wastewaters in VCE'S, and disposal of sludges and non-recyclable wastewaters in evaporation ponds. The process flow diagram and process and material balances for Option IV are shown on Figure 4-4 and Table 4-4.

Option IV uses minimal treatment of the raw water to the cooling towers, including sulfuric acid, scale inhibitor and chlorine. The cooling towers are operated at a maximum of 750 mg/l calcium and 150 mg/l silica. The resulting blowdown of 1,240 gpm is collected with the ash, FGD and miscellaneous wastewaters in a 5 million gallon surge pond for processing in VCE's. The FGD waste is directed to the evaporation ponds for disposal.

The total wastewater flow to the surge pond of 1,615 gpm is processed by four 500 gpm VCE's operating in parallel. The product flow of 1,551 gpm is more than sufficient to meet the total make-up needs of the ash, condensate and FGD systems. The resulting brine flow of 63 gpm is discharged to 31 acres of evaporation pond.

Option V - Make-Up Softening/VCE/Evaporation Ponds

Option V includes lime/soda softening of the total plant raw water supply, treatment of recyclable plant wastewaters in brine concentrator evaporators, and disposal of sludge and non-recyclable wastewaters in evaporation ponds. The process flow diagram and material balance for Option V are shown on Figure 4-5 and Table 4-5.

The raw water is processed by two 5,000 gpm softener/clarifiers operating in parallel at a total feed rate of 6,660 gpm. The effluent is controlled at 60 mg/l calcium and 11 mg/l of silica through the addition of lime and soda ash. The effluent is collected in a 2.5 million gallon collection basin for distribution to the station. Sulfuric acid, chlorine and scale inhibitor are added to the circulating water system to control scale and microbiological activity.

The cooling towers are operated at limits of 750 mg/l calcium and 150 mg/l silica. The resulting blowdown of 490 gpm is collected with the ash, condensate, FGD and miscellaneous wastewaters in a 5 million gallon surge pond for processing with VCE's.

The total wastewater inflow of 868 gpm to the surge pond is processed by two 500 gpm VCE's operating in parallel. The product production of 833 gpm satisfies the make-up needs of the ash, condensate and FGD system. The brine waste flow of 35 gpm and softener sludge of 10 gpm are transferred to a 24 acre evaporation pond system.

Summary of Equipment and Pond Capacities

The equipment and pond capacities for each option are summarized in Table 4-6.

TABLE 4-1 Ix Softening/Sidestream Silica/VTFE/Evaporation Ponds

Stream	gpm	Na	Ca	Mg	Cl	SO_4	HCO_3	SiO_2	TDS	SS
Make-Up (Credit Taken for Recovered Water)										
Palo Verde	6557	394	132	46	364	550	322	21	1830	10
Salt, lb/min		21.5	7.2	2.5	19.9	30.1	17.6	1.1	100.1	.5
Ion Exchange										
Feed	6091	394	132	46	364	550	322	21	1830	10
Salt, lb/min		20.0	6.7	2.3	18.5	27.9	16.4	1.1	92.9	.5
Softened Water	6091	518	40	30	364	550	322	21	1845	10
Salt, lb/min		26.3	2.0	1.5	18.5	27.9	16.4	1.1	93.7	.5
Ix Waste Eff	160	13765	600	1028	10542	19956	0	150	46041	20
Salt, lb/min		18.4	.8	1.4	14.1	26.7	0	2.0	61.6	.0
Ix Sludge	20	13765	600	1028	10542	19956	0	150	46041	100000
Salt, lb/min		2.3	.1	.2	1.7	3.3	0	.0	7.6	16.4
Cooling Towers at 29 Cycles of Concentration										
Evaporation	5900									
Drift	20	18305	640	374	10542	26891	80	150	56982	289
Salt, lb/min		3.1	.1	.1	1.8	4.5	.0	.0	9.5	.0
Blowdown	180	18305	640	374	10542	26891	80	150	56982	289
Salt, lb/min		27.5	1.0	.6	15.8	40.4	.1	.2	85.5	.4
SStream Feed	853	18305	640	374	10542	26891	80	150	56982	289
Salt, lb/min		130.2	4.6	2.7	75.0	191.3	.6	.8	405.4	2.1
SStream Out	842	19132	545	1	10542	26985	0	75	57280	10
Salt, lb/min		134.3	3.8	.0	74.0	189.5	0	.5	402.2	.1
SStream Sludge	11	19132	545	1	10542	26985	0	75	57280	100000
Salt, lb/min		1.8	.0	.0	1.0	2.5	0	.0	5.3	7.8
Chemical Feeds										
Caustic (50%)	17.5 lb/min 20.5 lb/kgal Sidestream Feed									
Sulfuric(93%)	15.9 lb/min 18.6 lb/kgal Sidestream Feed									
Soda Ash(98%)	2.0 lb/min 2.3 lb/kgal Sidestream Feed									
Wastewater To Treatment										
Ash System	100	400	350	200	400	1921	0	30	3301	150
Regen Waste	50	1970	660	230	1820	2750	0	105	7535	150
FGD Sludge	125	6677	500	200	3640	11000	0	250	22267	10000
Ix Waste Eff	160	13765	600	1028	10542	19956	0	150	46041	20
Miscellaneous	100	394	132	46	364	550	322	21	1830	10
Total to VTFE	535	6013	448	422	4320	9445	0	123	20649	20
		26.8	2.0	1.9	19.3	42.2	0	.5	92.2	.1
Product	485	5			7				12	0
Waste	51	52964	400	3717	38046	74673	0	200	170000	16308
Salt, lb/min		26.8	.2	1.9	19.3	37.9		.1	86.2	8.3
Final Disposal										
Ix Sludge	20	13765	600	1028	10542	19956	0	150	46041	100000
SStream Sludge	11	19132	545	1	10542	26985	0	75	57280	100000
VTFE Waste	51	52964	400	3717	38046	74673	0	200	170000	16308
Total	81	38903	468	2564	27671	54984	0	171	124761	47877

Evaporation Rate at 2 gpm/ac = 40.7 Acres of Disposal Ponds

TABLE 4-2 Sidestream Softening/VCE/Evaporation Ponds

Stream	gpm	Na	Ca	Mg	Cl	SO_4	HCO_3	SiO_2	TDS	SS
Plant Make-Up (Credit Taken for Recovered Water)										
Palo Verde	6701	394	132	46	364	550	322	21	1830	10
Cooling Towers at 28 Cycles of Concentration										
Make-Up										
Palo Verde	6121	394	132	46	364	550	322	21	1830	10
Salt, lb/min		20.1	6.7	2.3	18.5	28.0	16.4	1.1	93.2	.5
Evaporation	5900									
Drift	20	16909	600	209	10072	23870	80	150	51891	292
Salt, lb/min		2.8	.1	.0	1.7	4.0	.0	.0	8.7	.1
Blowdown	180	16909	600	209	10072	23870	80	150	51891	292
Salt, lb/min		25.4	.9	.3	15.1	35.8	.1	.2	77.9	.5
SStream Feed	1368	16909	600	209	10072	23870	80	150	51891	292
Salt, lb/min		192.9	6.8	2.4	114.9	272.2	.4	1.7	591.8	3.5
SStream Out	1346	17881	100	21	10072	24020	0	75	52131	10
Salt, lb/min		200.8	1.1	.2	113.1	269.7	0	.8	585.3	.1
SStream Sludge	21	17881	100	21	10072	24020	0	75	52131	150000
Salt, lb/min		3.2	.02	.00	1.8	4.2	0	.01	9.2	26.5
Chemical Feeds										
Lime (85%)	6.3	lb/min	4.6	lb/kgal	Sidestream Feed					
Soda Ash(98%)	25.8	lb/min	18.8	lb/kgal	Sidestream Feed					
Sulfuric(93%)	16.0	lb/min	11.7	lb/kgal	Sidestream Feed					
Wastewater										
To Treatment										
Ash System	100	400	350	200	400	1921	0	30	3301	150
Regen Waste	50	1970	660	230	1820	2750	0	105	7535	150
Cooling Towers	180	16909	600	209	10072	23870	80	150	51891	292
FGD Sludge	125	6677	500	200	3640	11000	0	250	22267	10000
Miscellaneous	100	394	132	46	364	550	322	21	1830	10
Total to BC	555	7308	454	178	4388	11108	0	124	23559	20
Salt, lb/min		33.8	2.1	.8	20.3	51.4	0	.6	109.0	.1
Product	495	2							5	0
Waste	60	56228	400	1369	33760	78043	0	200	170000	14188
Final Disposal										
SStream Sludge	21	17881	100	21	10072	24020	0	76	52131	150000
BC Waste	60	56288	400	1369	33760	78043	0	200	170000	14188
Total	81	46232	322	1017	27585	63961	0	157	139275	49591

Evaporation Rate at 2 gpm/ac = 40.7 Acres of Disposal Ponds

TABLE 4-3 Evaporation Ponds

Stream	gpm	Na	Ca	Mg	Cl	SO_4	HCO_3	SiO_2	TDS	SS
Make-Up (Credit Taken for Recovered Water)										
Palo Verde	8313	394	132	46	364	550	322	21	1830	10
Salt, lb/min		27.3	9.2	3.2	25.2	38.1	22.3	1.5	126.9	.7
93% Sulfuric	2.2 lb/kgal Tower Make-Up									
Cooling Towers	5.7 Cycles of Concentration									
Evaporation	5900									
Drift	20	2234	748	261	2064	4483	100	119	10377	57
Salt, lb/min		.4	.1	.0	.3	.7	.0	.0	1.7	.0
Blowdown	1243	2234	748	261	2064	4483	100	119	10377	57
Salt, lb/min		23.2	7.8	2.7	21.4	46.5	1.0	1.2	107.6	.6
Wastewater To Ponds										
Ash System	100	400	350	200	400	1921	0	30	3301	150
Regen Waste	125	1970	660	230	1820	2750	0	105	7535	150
Cooling Towers	1243	2234	748	261	2064	4483	100	119	10377	57
FGD Sludge	125	6677	500	200	3640	11000	0	250	22267	10000
Miscellaneous	100	394	132	46	364	550	322	21	1830	10
Total to Ponds	1693	2326	664	238	1964	4452	92	117	10123	800
Salt, lb/min		32.8	9.4	3.4	27.7	62.9	1.3	1.6	143.0	11.3

Evaporation Rate at 2 gpm/ac = 844.6 Acres of Disposal Ponds

TABLE 4-4 VCE/Evaporation Ponds

Stream	gpm	Na	Ca	Mg	Cl	SO_4	HCO_3	SiO_2	TDS	SS
Make-Up (Credit Taken for Recovered Water)										
Palo Verde	6685	394	132	46	364	550	322	21	1830	10
Salt, lb/min		22.0	7.4	2.6	20.3	30.7	18.0	1.2	102.0	.6
93% Sulfuric	2.2 lb/kgal Tower Make-Up									
Cooling Towers	5.7 Cycles of Concentration									
Evaporation	5900									
Drift	20	2239	750	261	2068	4475	100	119	10398	57
Salt, lb/min		.4	.1	.0	.3	.7	.0	.0	1.7	.0
Blowdown	1240	2239	750	261	2068	4475	100	119	10398	57
Salt, lb/min		23.2	7.8	2.7	21.4	46.3	1.0	1.2	107.6	.6
Wastewater To Treatment										
Ash System	100	400	350	200	400	1921	0	30	3301	150
Regen Waste	50	1970	660	230	1820	2750	0	105	7535	150
Cooling Towers	1240	2239	750	261	2068	4475	100	119	10398	57
FGD Sludge	125	6677	500	200	3640	11000	0	250	22267	10000
Miscellaneous	100	394	132	46	364	550	322	21	1830	10
Total to BC	1615	2346	665	239	1973	4761	0	117	10258	20
Salt, lb/min		31.6	9.0	3.2	26.6	64.1	0	1.6	138.2	.3
Product	1553	2							5	0
Waste	62	48909	400	4973	41144	66965	0	200	162592	49995
Final Disposal										
BC Waste	62	48909	400	4973	41144	66965	0	200	162592	49995

Evaporation Rate at 2 gpm/ac = 31.0 Acres of Disposal Ponds

TABLE 4-5 Make-Up Softening/VCE/Evaporation Ponds

Stream	gpm	Na	Ca	Mg	Cl	SO_4	HCO_3	SiO_2	TDS	SS
Make-Up (Credited Taken for Recovered Water)										
Palo Verde	6664	394	132	46	364	550	322	21	1830	10
Salt, lb/min		21.9	7.3	2.6	20.2	30.6	17.9	1.2	101.7	.6
Make-Up Softening										
Feed	6664	394	132	46	364	550	322	21	1830	10
Salt, lb/min		21.9	7.3	2.6	20.2	30.6	17.9	1.2	101.7	.6
85% Lime	3.8	lb/kgal Softener Feed								
98% Soda Ash	3.2	lb/kgal Softener Feed								
93% Sulfuric	2.5	lb/kgal Softener Feed								
Effluent	6655	554	60	5	364	825	0	10	1818	10
Salt, lb/min		30.7	3.3	.3	20.2	45.8	0	.6	100.9	.6
Sludge	10	554	60	5	364	825	0	10	1818	200000
Salt, lb/min		.05	.01	.00	.04	.08	0	.00	.18	16.1
Cooling Towers at 12.5 Cycles of Concentration										
Evaporation	5900									
Drift	20	6919	750	58	4550	10314	0	131	22722	125
Salt, lb/min		1.2	.1	.0	.8	1.7	0	.0	3.8	.0
Blowdown	493	6919	750	58	4550	10314	0	131	22722	125
Salt, lb/min		28.5	3.1	.2	18.7	42.4	0	.5	93.4	.5
Wastewater To Treatment										
Ash System	100	400	350	200	400	1921	0	30	3301	150
Regen Waste	50	1970	660	230	1820	2750	0	105	7535	150
Cooling Towers	493	6919	750	58	4550	10314	0	131	22722	125
FGD Sludge	125	6677	500	200	3640	11000	0	250	22267	10000
Miscellaneous	100	394	132	46	364	550	322	21	1830	10
Total to BC	868	5097	592	103	3301	7998	0	122	17213	20
Salt, lb/min		21.0	2.4	.4	13.6	32.9	0	.5	70.8	.1
Product	830	2							5	0
Waste	38	60357	400	1220	39097	78859	0	200	180134	25506
Final Disposal										
Softener Sludge	10	554	60	5	364	825	0	10	1818	200000
BC Waste	38	60357	400	1220	39097	78859	0	200	180134	25506
Total	48	47347	326	956	30671	61884	0	159	141343	63465
Evaporation Rate at 2 gpm/ac = 24.0 Acres of Disposal Ponds										

TABLE 4-6

Equipment and Pond Capacities

	Options				
Description	I	II	III	IV	V
Ion Exchange Capacity, gpm	10,000				
VTFE Capacity, gpm	700				
Sidestream Softener Capacity, gpm	1,200	2,000			
Collection Basin, mgal	2.5	2.0			
Surge Ponds, mgal	6.0	7.0		5.0	5.0
Raw Water Softeners					10,000
VCE Capacity, gpm		700		2,000	1,000
Evaporation Ponds, Acres	41	41	845	31	24

FIGURE 4-1

OPTION I PROCESS FLOW DIAGRAM

STREAM	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26	27	28	29
Flow, gpm (lb/min)	6557	6091	6091	20	3900	853	11	842	100	150	20	990000	33600	steam	300	30	100	120	100	535	485	485	50						
pH	8.0	S		8.0		S	11.0	8.0	S	8.0	8.0	8.0	8.0		9.5	2-14	12	8.0	S	5-6		5-6	5-6						
Sodium mg/l CO	394	A	518	18305		A	19137	S	A	13765	S	S	S		5	1940	400	6677	A	6813		5	52964						
Calcium	132	M	42	640		M	545	T	M	600	A	A	A			660	350	500	M	447			400						
Magnesium	46	E	30	374		E	1	R	E	1028	M	M	M			230	200	200	E	622			3717						
Ammonia		A				A		A	A		E	E	E						A										
Nitrate	1	S	1			S		S	S		A	S	S						S										
Chloride	364	S	364	10542		S	10542	S	S	10542	S	S	S		7	1828	400	3640	S	4328		7	28946						
Bicarbonate	322	T	322	80		T		T	T		T	T	T						T										
Carbonate		R				R		E	R		E	E	E						R										
Sulfate	330	E	330	26891		E	26891	A	E	19956	A	A	A			2750	1921	11000	E	9445			74673						
Phosphate	0.1	A	0.1			A		M	A		M	M	M						A										
Silica	21	M	21	150		M	75	7	M	150	10	4	4			105	30	250	M	122			200						
Dissolved Solids	1830	1	1845	56986		4	57284		4	46041					12	7535	3301	22267	1	20649		12	170000						
Suspended Solids	10		10	294			QD	10	294	QD	20	294	294			150	150	10000		20			16308						

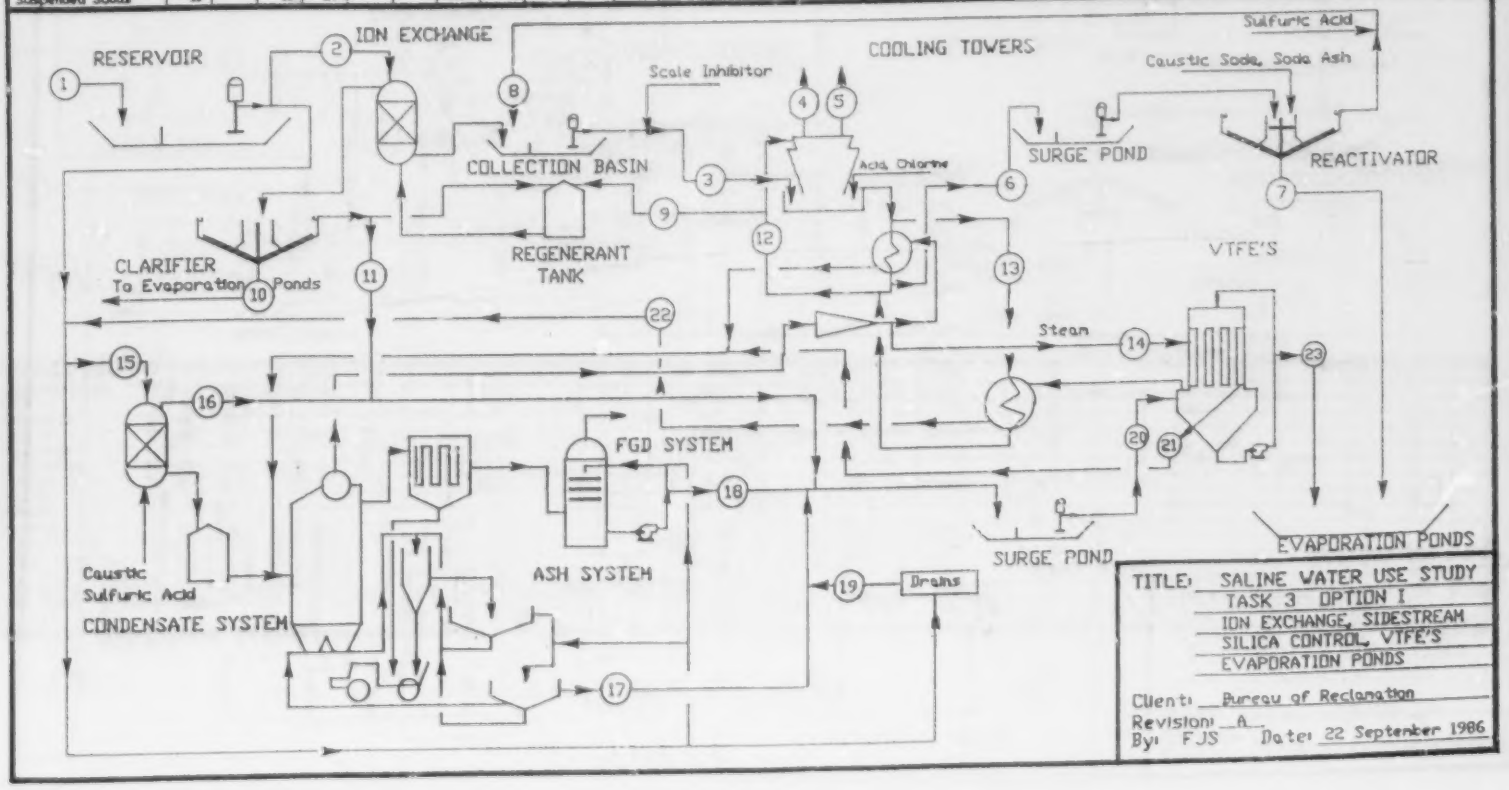


FIGURE 4-2

OPTION II PROCESS FLOW DIAGRAM

STREAM	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26	27	28	29
Flow, gpm (lb/min)	6701	580	6121	20	3900	180	1368	.21	1346	390000	300	56	125	100	100	555	495	60											
pH	8.0	8.0	8.0	8.0		8.0	8.0	11	8.0	S	5-6	2-14	6	S	10	5-6	5-6	5-6											
Sodium mg/l CO	394	394	394	16909		S	S	17891	+	A	5	1970	6677	A	400	7308	5	56226											
Calcium	132	132	132	600		A	A	100	A	M		660	500	M	350	454		498											
Magnesium	46	46	46	289		M	M	21	L	C		230	184	C	200	178		1569											
Ammonia						E	E		A																				
Nitrate	1	1	1			A	A		I	A				A															
Chloride	364	364	364	10072		S	S	10072	S	S	7	182	3648	S	400	4368	7	33768											
Bicarbonate	322	322	322	89		T	T		I	T				T															
Carbonate						R	R		A	R				R															
Sulfate	550	550	550	2887		E	E	2400	N	E		2750	11000	E	1921	11109		78043											
Phosphate	0.1	0.1	0.1			A	A		E	A				A															
Silica	21	21	21	150		M	M	73		M		105	250	M	30	124		209											
Dissolved Solids	1830	1830	1830	31891		A	A	52132			12	7535	22267	S	3911	23539	12	179908											
Suspended Solids	20	10	10	300				0.5	10			150	10000		150	20		14288											

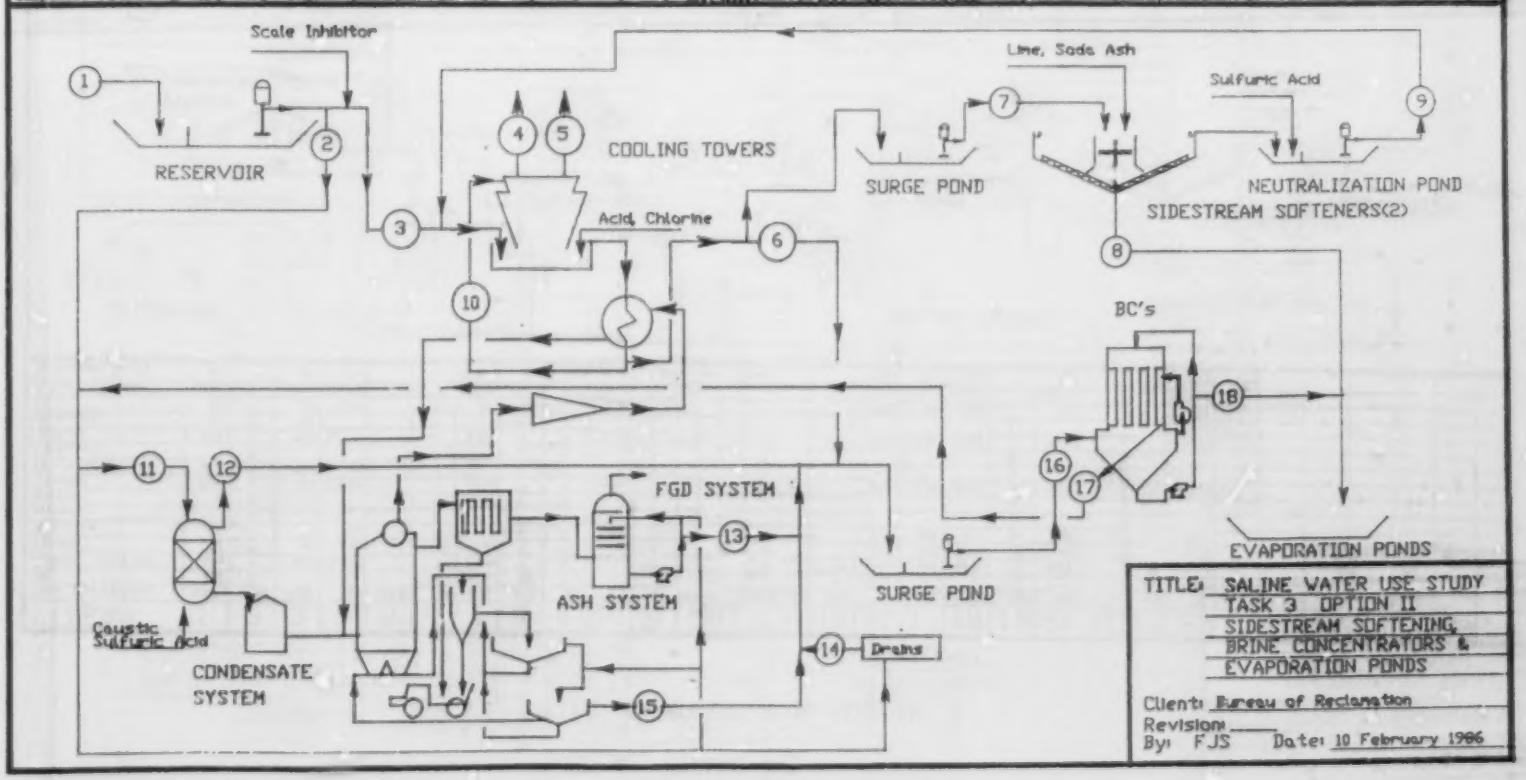


FIGURE 4-3

OPTION III PROCESS FLOW DIAGRAM

STREAM	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26	27	28	29
Flow, gpm (lb/hr)	6313	7163	29	5920	1150	390000	1243	300	125	125	100	100																	
pH	8.0	8.0	8		S	S	S	S	2-14	6	10	S																	
Sodium mg/l CO	394	394	2224		A	A	A	A	1970	6677	400	A																	
Calcium	132	132	730		M	M	M	M	660	500	350	M																	
Magnesium	46	46	261		E	E	E	E	230	200	200	E																	
Ammonia					A	A	A	A				A																	
Nitrate	1	1	6		S	S	S	S				S																	
Chloride	364	364	2064		S	S	S	S	1820	3640	440	S																	
Bicarbonate	322	322	80		T	T	T	T				T																	
Carbonate					R	R	R	R				R																	
Sulfate	330	330	4483		E	E	E	E	2750	11000	1921	E																	
Phosphate	0.1	0.1	0.6		A	A	A	A				A																	
Silica	21	21	119		M	M	M	M	105	238	30	M																	
Dissolved Solids	1830	1830	10377		2	3	3	2	7335	22267	3001	2																	
Suspended Solids	20	10	57						150	10000	150																		

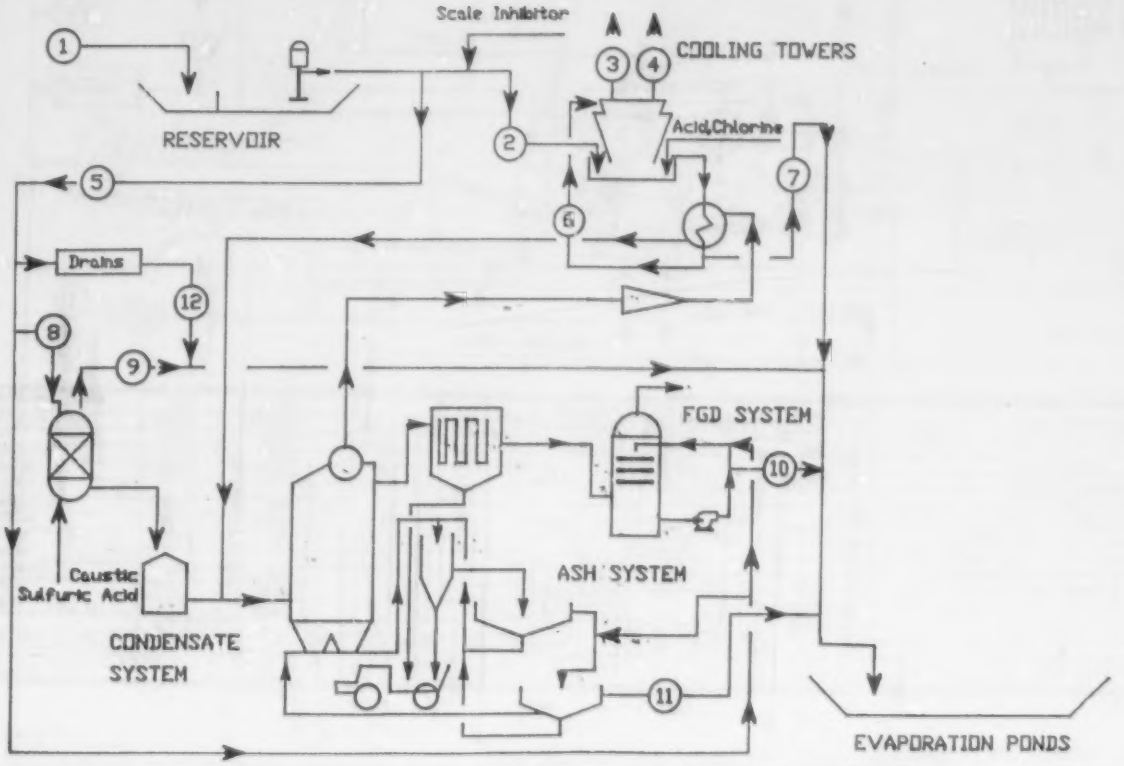


FIGURE 4-4

OPTION IV PROCESS FLOW DIAGRAM

STREAM	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26	27	28	29
Flow, gpm Qb/mn	6685	7160	20	5900	1076	390000	1240	300	50	125	100	1615	1533	63	100														
pH	8.0	S	8		5-6	S	S	S	2-14	6	10	5-6	5-6	5-6	S														
Sodium mg/l CO	394	A	2239		3	A	A	A	1970	6677	466	2346	3	48909	A														
Calcium	132	M	750			M	M	M	650	500	350	665	A	400	M														
Magnesium	46	L	261			E	E	E	230	200	200	239	M	4973	L														
Ammonia						A	A	A					E																
Nitrate	1	A	6			S	S	S					A		A														
Chloride	364	S	2068		7	S	S	S	1820	3640	400	1973	S	41144	S														
Bicarbonate	322	T	100			T	T	T					S		T														
Carbonate		R				R	R	R					T		R														
Sulfate	550	E	4475			E	E	E	2750	11000	1921	4761	R	66965	E														
Phosphate	0.1	A	0.6			A	A	A					E		A														
Silica	21	M	119			M	M	M	105	256	36	117	A	200	M														
Dissolved Solids	1830	1	10390		12	3	3	3	7535	22267	3301	10258	M	170000	1														
Suspended Solids	20		57						150	10000	150	20	3	50000														

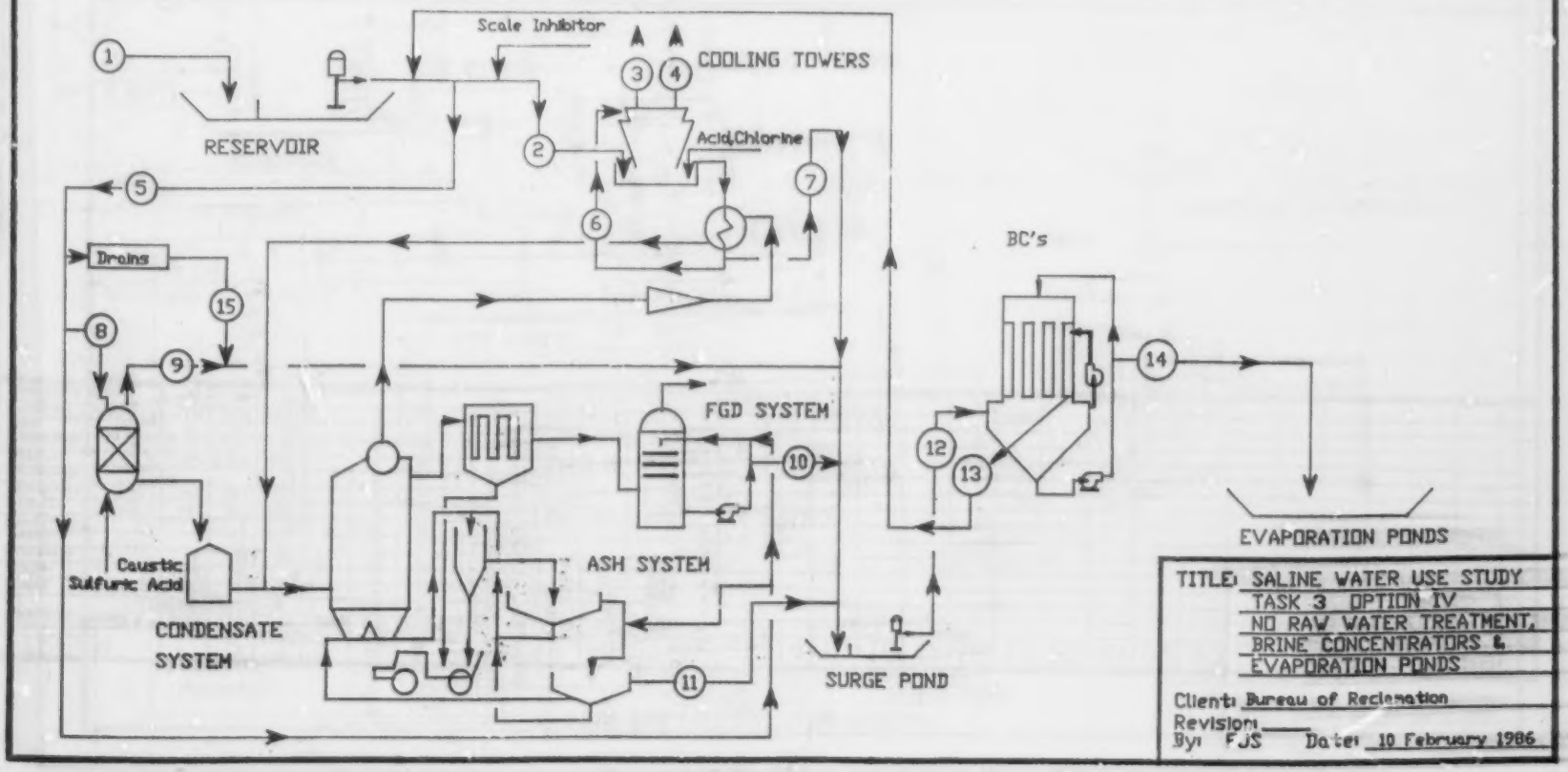
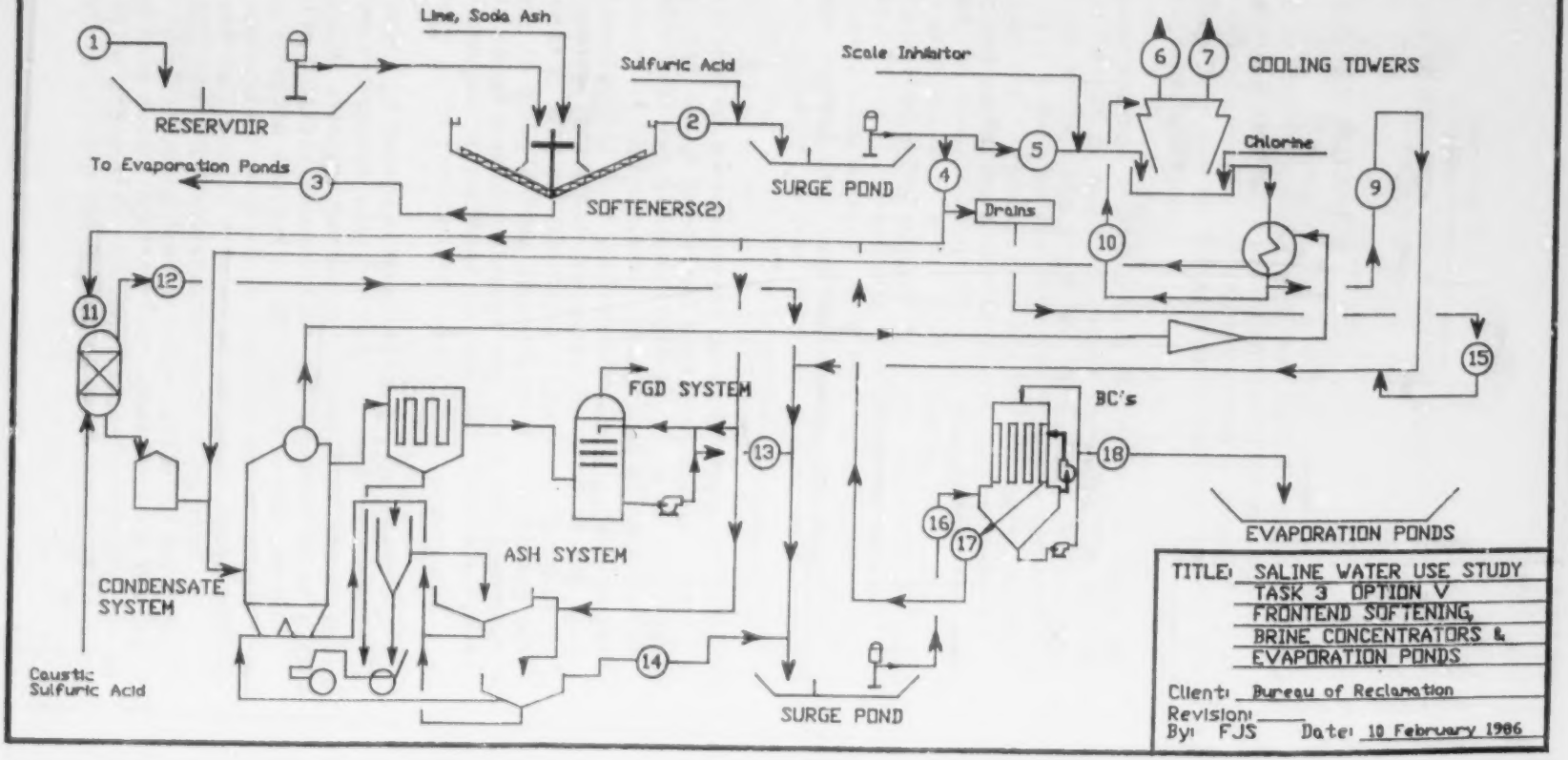


FIGURE 4-5

OPTION V PROCESS FLOW DIAGRAM

STREAM	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26	27	28	29
Flow, gpm (lb/hr)	6664	6654	10	251	6413	20	5700		493	720000	200	38	125	188	188	868	833	35											
pH	8.0	10.5	10.5	7	7	7			7	3	9-6	8-14	6	10	3	5-6	5-6	5-6											
Sodium mg/l CO	394	554	554	554	554	6919			6919	A	5	1970	6677	400	A	5897	5	68357											
Calcium	132	60	60	60	60	730			730	M		660	500	350	M	592		498											
Magnesium	46	5	5	5	5	58			58	E		230	200	200	E	183		1280											
Ammonia										A					A													
Nitrate	1									S					S													
Chloride	364	364	364	364	364	4550			4550		7	1820	3640	400	S	3301	7	39997											
Bicarbonate	328									O					O													
Carbonate										T					T													
Sulfate	550	550	550	825	825	10314			10314	A		2750	11000	1921	E	7999		78835											
Phosphate	0.1									A					A													
Silica	21	10	10	10	10	131			131	M		105	250	2*	M	122		289											
Dissolved Solids	1838	1543	1543	1818	1818	22722			22722	9	12	7535	22267	3301	1	17213	12	172000											
Suspended Solids	20	10	ORD	10	10	125			125			150	10000	150		20		25306											



Section 5

APPLICATION OF ETIWANDA FIELD TEST RESULTS

DESCRIPTION OF ETIWANDA FIELD TEST PROGRAM

A simplified flow schematic of the Etiwanda Field Test Unit is shown in Figure 5-1. The feed water to the field test unit was a synthesis of average effluent from the Palo Verde Outfall drain. It was made from a combination of Colorado River water, blowdown from one of the station's cooling towers, and sodium chloride mixed in proper proportion.

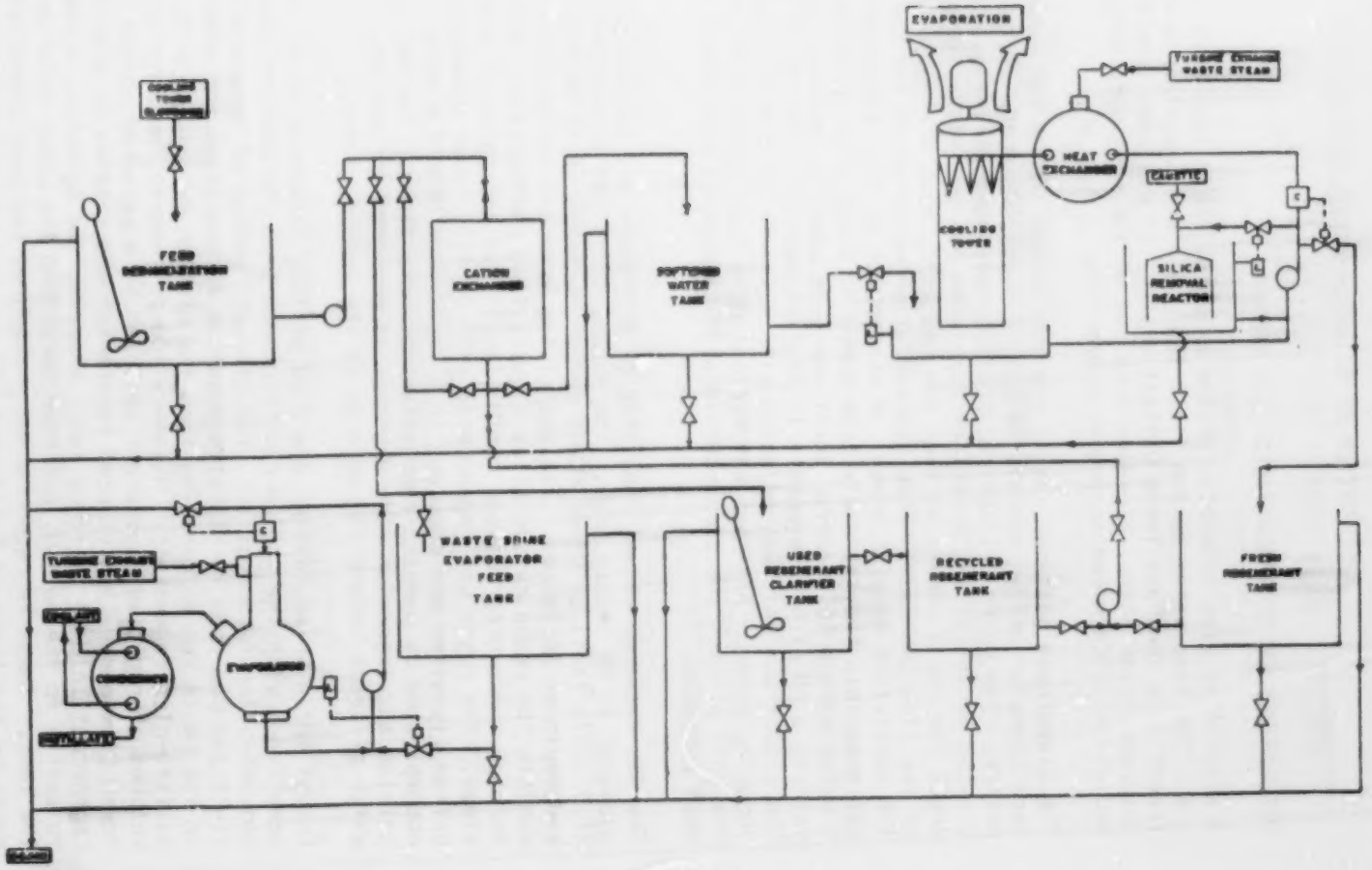
The sidestream softener vessel was fabricated from a FRP vessel having a 48-inch diameter and an overall height of about 65 inches including the conical bottom. The upper cylindrical portion was about 48 inches high and was subdivided by installing a central plexiglas box and plates to provide separate areas for rapid mixing of feed and reagents, slow mixing, and axial downflow of the floc and feed. Floc sedimentation occurred in the conical base and in the subdivided annular areas. A suspension of floc was pumped from the conical area into the rapid mix zone where it mixed with incoming water and treated with a 10% caustic solution. The caustic feed was regulated by maintaining a pH of approximately 11 in this section. The pH of the effluent from the sidestream unit was adjusted with sulfuric acid to a value equal to the incoming water. Although not typical of a full scale unit, the segmenting of the unit allowed the cross-sectional rise area of the vessel to varied for test purposes.

The ion exchange vessel consisted of a transparent acrylic column 10 inches in diameter by 96 inches high. The vessel contained approximately 12.7 gallons (1.5 cubic feet) of Dowex HCR-S cation exchange resin having a settled resin bed depth of 36 inches. The vessel had screened connections at each end to contain the resin but to allow liquid, fine particulates, and gypsum to pass through. A microprocessor controlled valves and pumps to sequence the Ix steps. The resin was regenerated with cooling tower blowdown both in an upflow fluidized bed mode and in a rapid-flow downflow mode. Softening was accomplished by pumping feed water downward through the resin bed such that calcium and magnesium were absorbed in exchange for sodium. The Ix softened water effluent became the make up to the cooling tower.

Except for a brief initial start up period, regeneration of the resin was accomplished with the sodium-rich blowdown from the cooling tower operated at a concentration factor of 30. The standard method of regeneration was an upflow fluidized bed mode to minimize possible accumulation and cementation of gypsum in the resin bed originating from the elution of calcium with the sodium sulfate-rich regenerant. Regeneration procedures included use of about 10 bed volumes of recycled regenerant followed by a balanced volume of fresh cooling tower blowdown. The recycled regenerant was created by placing used regenerant into an agitated tank, stopping the agitator, allowing the gypsum to settle to the bottom, and then decanting the clear solution from the top of the tank. These Ix regeneration procedures had been investigated in previous tests by the State of California [7] and the Bureau of Reclamation [5][6]. A

FIGURE 5-1

FLOW SCHEMATIC-ETIWANDA FIELD TEST UNIT



second method of regeneration tested at Etiwanda used the same regenerant solutions just described but was accomplished in a downflow mode through a packed resin bed utilizing a higher flow rate. This second method attempted to better utilize the resin through a lowering of the regeneration time. If successful, this approach could lower the Ix capital requirement of a full size plant provided that other factors remained the same.

EVALUATION OF FIELD TEST RESULTS

Ion Exchange System

A total of 12 Ix cycles were completed using the fluidized bed upflow mode and 7 cycles using the downflow packed bed mode. The number of Ix cycles completed during 2 months of testing was limited by the evaporation rate in the pilot plant cooling tower. Representative Ix performance had been achieved prior to completion of the last cycle of each mode. Thus, the Ix data collected during these cycles were sufficient to accurately define Ix performance of these two regeneration modes under the conditions operated. It is noteworthy that no precipitation of gypsum in the resin bed was observed during either regeneration mode because the test water contained precipitation inhibitors routinely added to the Etiwanda station's cooling towers.

The Ix performance of the two regeneration modes was virtually identical in terms of projected resin requirements of a full size plant. While the downflow packed bed mode did result in a lower regeneration and cycle time as expected, the volume of water softened per cycle as reflected in the calcium resin capacity was also proportionally less. The reasons attributed for the lower resin capacity of the downflow mode tested are the smaller contact time allowed for regeneration because of the higher flow rate with the same volume of regenerant plus the fact that softening and regeneration were accomplished in the same flow direction (coflow). With the fluidized bed mode the softening and regeneration occurred in opposite directions (counterflow). Numerous studies have demonstrated Ix performance advantages of counterflow as compared to coflow.

Regeneration via the packed bed downflow mode was not included in the present process feasibility analysis because it showed no performance advantage but at the same time had the following additional disadvantages as compared to the fluidized bed mode:

- An additional water stream would be required for a backwash prior to the downflow packed bed regeneration, which would affect the overall plant water balance and require extra equipment and procedures. In the case of fluidized bed regeneration, the backwash is already included as the first part of the regeneration step using recycled regenerant.

- The cooling tower blowdown brine may have to be filtered prior to use as regenerant through a packed resin bed if the suspended solids present create an excessive pressure drop. Suspended solids can enter the brine as windblown dust to the cooling tower and be concentrated in the circulating water system. The extra filtration

step would add to capital and operating costs and add an waste stream. The fluidized bed regeneration mode by its very nature is not affected by suspended solids in the regenerant.

- The higher regeneration flow rates through a packed bed may require larger capacity pumps, higher pressure ratings on piping and vessels, and greater energy consumption, which would all increase costs.

- The ability of the packed bed mode to handle gypsum precipitation in the resin bed has not been demonstrated in the absence of scale inhibitor. It cannot be assured that scale inhibitor would always be present in the cooling tower blowdown. Further testing would be required to establish whether gypsum precipitation could be handled safely during coflow regeneration through a packed bed when scale inhibitor is absent. On the other hand, successful fluidized bed regeneration under gypsum precipitation conditions has been demonstrated repeatedly in field test programs.

The Ix performance during the Etiwanda pilot plant tests was approximately twice as high as pretest assumptions used in the preliminary feasibility calculations. A pretest assumption was that 85 bed volumes could be softened per ix cycle. However, the Etiwanda pilot plant achieved 175 bed volumes, corresponding to a calcium resin capacity of 1.1 meq/l of resin. These pretest assumptions had been based on projections of laboratory equilibrium data with a conservative factor included to account for resin-bed breakthrough dynamics and other uncertainties. The softening flow rate tested was also considerably greater than assumed earlier in order to try to lower projected Ix capital costs. The pretest assumption was a conservative 2.2 gpm/ft3, but the loading rate was increased to 4.6 gpm/ft3 during the tests. As a result, the pilot plant cycle time to accomplish the softening and regeneration performance stated here was 6 hours.

The Ix was intentionally operated so as to allow some magnesium breakthrough as needed by the sidestream softener for silica removal downstream. An average 20 mg/L of magnesium, or 42 percent of the influent, was allowed to pass up to the point of softening termination. At the same time 95 percent of the calcium was removed down to an average of 6.6 mg/l in the effluent. Had more water been allowed to be softened prior to termination of each Ix softening step an even greater proportion of magnesium would have been passed. The overall Ix performance would have also increased slightly with a modest increase in calcium leakage. However, such performance fine tuning was beyond the scope of the Etiwanda testing and the needs of the present feasibility analysis.

Sidestream Softener

Operating results from the sidestream softener at the Etiwanda field test program for the period November 24 through December 30, 1986, were evaluated for the following:

- The accuracy of the design basis used in the economic comparison of the process options.

- The impact of the sidestream softener on the ion exchange system through its role in modifying the chemistry of the circulating water which was used for regeneration of the ion exchange system.

For systems similar to the Etiwanda field test program, the primary effect of the sidestream softener unit on the ion exchange system is the modification of the circulating water which is used for ion exchange regeneration. Although the ion exchange system is used to remove hardness (calcium and magnesium), the unit must be operated such that sufficient magnesium passes through to allow precipitation of magnesium hydroxide in the sidestream unit. Dissolved silica in the circulating water is then adsorbed onto the precipitated magnesium hydroxide. During this process, the total and relative concentrations of sodium, calcium, and magnesium concentrations are altered. Also any sulfuric acid added to stabilize the softener effluent increases the sulfate concentration in the circulating water. In general, the regenerant should have a high concentration of sodium and low concentrations of calcium, magnesium, and sulfate. The presence of calcium and magnesium tend to reduce resin capacity whereas high concentrations of sulfate can lead to precipitation of calcium sulfate in the resin bed during regeneration.

Flow from the circulating water loop was directed to the sidestream softener reactor where sodium hydroxide was added. The effluent discharged to a neutralization tank where sulfuric acid was added. This stream was returned to the circulating water loop. The added caustic soda induced three reactions:

- Precipitation of magnesium hydroxide upon which silica was absorbed

- Conversation of bicarbonates to carbonates which in turn reacted with calcium to form solid $CaCO_3$

- Reaction with dissolved CO_2 to form additional carbonates and in turn $CaCO_3$ solids

The pH was controlled to a value of about 11.0. In the process, silica, magnesium, and calcium were removed from the side stream flow as solid phase materials, which settled in the clarifier portion of the reactor.

The field test results and comparative sidestream silica control system operating conditions are shown in Tables 5-1 and 5-2. As shown in Table 5-2, the reactor was operated at about a 200-fold smaller flow rate than would be encountered in a conventional system. The Etiwanda field test program, however, achieved near normal silica reduction as a result of being operated at about 1/35th of its normal flow rate for 19% of the time. Another feature of the operation was that sodium hydroxide chemical feed was almost double that found necessary in jar tests. Likewise, it was about double the concentration projected for the economic analysis. Also, the rapid mix device and sludge

TABLE 5-1

ETIWANDA FIELD TEST RESULTS - SIDESTREAM SILICA CONTROL SYSTEM

DATE	REGEN (gal)	Qs (gpm)	HRS RUN (hrs)	SS VOL (gal)	Qm (gal)	SiO2 circ	SiO2 ss,ef	Mg ix,ef	Mg circ	Mg ss,ef	Ca ix,ef	Ca circ	Ca ss,ef
14-Nov-86	65				2272			35.2					
19-Nov-86	80				2168			28			32	88.2	24.1
21-Nov-86	64		244		2267						18		
----------		Time Period Considered In											
24-Nov-86		0.25	12	180		66	21		189.6	48.6			
25-Nov-86	75	0.23	11	151.8	3235	76	20		286.6	49.8		78.2	26.1
26-Nov-86		0.25	9	135									
01-Dec-86		0.23	9.5	131.1									
02-Dec-86	80	0.21	10.5	132.3	2265			34			12		
03-Dec-86		0.21	10	126		71	22.5		142	69.3		46	27
04-Dec-86	80	0.18	10.5	113.4	2414				142	91.5			
05-Dec-86		0.2	10.5	126									
10-Dec-86	80	0.25	4	60	2379			20.7			8		
11-Dec-86		0.21	6.5	81.9									
12-Dec-86	75	0.16	10.5	100.8	2283	72.5	29	26.7			16		
15-Dec-86		0.16	10.5	100.8									
16-Dec-86	60	0.155	10.5	97.65	2273						24		
17-Dec-86		0.2	10.5	126									
18-Dec-86		0.2	10.5	126									
19-Dec-86	80	0.16	10.5	100.8	1986								
22-Dec-86	80	0.205	11	135.3	2020			25.5					
29-Dec-86	80	0.18	9	97.2	629.1	88	18		257.6	113	14	64.1	25.1
30-Dec-86		0.18	8.5	91.8									
Sum	690		173	2034	19484								
Mean		0.201				74.7	22.1	26.7	203.6	74.4	14.8	62.8	26.1

recirculation were left on continuously 24 hours a day throughout the test period even though the reactor was out of service.

TABLE 5-2

SIDESTREAM SILICA CONTROL SYSTEM OPERATING CONDITIONS

MAKEUP ION EXCHANGE SYSTEM	
Volume of Effluent (gal)	19484
Average Flow Rate (gpm)	0.366
Total Volume of Regenerate (gal)	690
SIDESTREAM SILICA SYSTEM	
Volume Processed (gal)	2034
Flow Rates (gpm)	
Average for Test Period	0.038
Average while Operated	0.201
Normal Rate	7.000
Reactor	
Volume (gal)	420
Residence Time (min)	
Average for Test Period	11002
While Operated	2090
Normal Rate	60
NaOH Feed (mg/l)	
While Operated	1008
Jar Test Projections	581
Economic Design Basis	590
SYSTEM RATIOS	
This Test (Qs/Qm)	0.104
Economic Design (Qs/Qm)	0.104
Full Scale System	
Qs, Design Basis (gpm)	853
Qs, Projected from This Test (gpm)	634

This could result in CO2 absorption from the atmosphere into this strongly agitated vessel at a pH of 11 (i.e., an acid reacting with a base.)

The extremely large flow capacity of the sidestream softener system relative to the ion exchange system would suggest that the sidestream system distorted the chemistry for regeneration. However, the sidestream softening system was operated only 19% of the time and at 1/35th of its rated capacity. The resulting circulating water chemistry was reasonably similar to what would be expected in a full scale system operating under "normal" conditions. Thus, impact on ion exchange regeneration by the sidestream softening system appears to be small.

Mass balance calculations for softener performance are shown in Table 5.3, which includes three critical constituents--silica, calcium, and magnesium. This illustrates the system and the boundaries drawn around it for purposes of the mass balance calculation. In particular, the focus was on inputs and outputs of silica, calcium, and magnesium around the circulating water system loop. If the flow and chemical concentration measurements were error free and all streams were accounted for including all changes from the soluble phase, then the mass balance errors would approach zero. A positive mass difference value in the table indicates material which was unaccounted for (more material going in than coming out), whereas a negative value indicates more material out than in. Positive values could be attributed to one or more of the following factors:

- Errors in flow or concentration measurement
- Unaccounted for input or output streams
- A constituent being precipitated in the circulating water system

Because the volumes and concentrations were a "snapshot" taken in the middle of the operation, they did not reflect startup or shutdown changes, and as such should be a reasonable representation of the mass balance.

The results for magnesium are within the error margins that might be expected in such a system. For silica and calcium, 39% and 50% respectively of the mass is unaccounted for. Without additional data, it is not possible to draw definite conclusions for the discrepancies. Assuming a tight system and good analytical control, one possible explanation for the missing calcium might be the occurrence of precipitation with the circulating water loop, which was maintained at pH above 9. Another explanation might be related to CO2 adsorption and resultant CaCO3 precipitation in the ion exchange softener that was not detected.

Of greater concern is the disappearance of silica apparently somewhere in the system other than the sidestream softener. If this is true, the two most likely possibilities are (1) incorporation within the biological slime that had accumulated within the system (which was noted to be quite severe) or (2) chemical scale within the system. The effect of such a disappearance is that

TABLE 5-3

SIDESTREAM SILICA CONTROL SYSTEM - MASS BALANCE RESULTS

	Total Vol Processed (gal)	SiO2 (mg/l)	SiO2 (lb)	Ca (mg/l)	Ca (lb)	Mg (mg/l)	Mg (lb)
System Inputs							
Ion Exchange Effluent	19484	14.8	2.40	14.8	2.40	26.7	4.34
Sidesoft Effluent	2034	22.1	0.37	26.1	0.44	74.4	1.26
Total			2.78		2.85		5.60
System Outputs							
Sidesoft Influent	2034	74.7	1.27	62.8	1.07	257.6	4.37
Ion Exchange Regen	690	74.7	0.43	62.8	0.36	257.6	1.48
Total			1.70		1.43		5.85
Mass Difference (In/Out) %			39		50		-4

it places less of a demand on the performance of the sidestream silica control system (by 39%) and/or casts doubt on whether scale may be a problem in such a system. Longer term tests involving carefully planned balance and accounting of all of the silica would be necessary to determine the exact cause.

Even neglecting the apparent 39% lower silica that must be controlled in the system, the test results suggested that the softener size was 40% larger than the results experienced with a full scale system. This indicated larger size combined with a necessity to deal with the unaccounted "disappearing" silica, suggest an even larger discrepancy from noted experience.

Dissolved silica is removed from solution in a lime or caustic softener by adsorption onto magnesium hydroxide solid surfaces. The surface may originate from dissolved magnesium precipitated from solution, or solid particles of added magnesium oxide (MgO). Results from a number of operating industrial facilities, pilot facilities, and limited laboratory results are shown in Figure 5-2. The data were evaluated as an adsorption isotherm relative log10 (mg/l SiO_2 removed per mg/l soluble magnesium removed) vs. log10 (mg/l effluent SiO_2.)

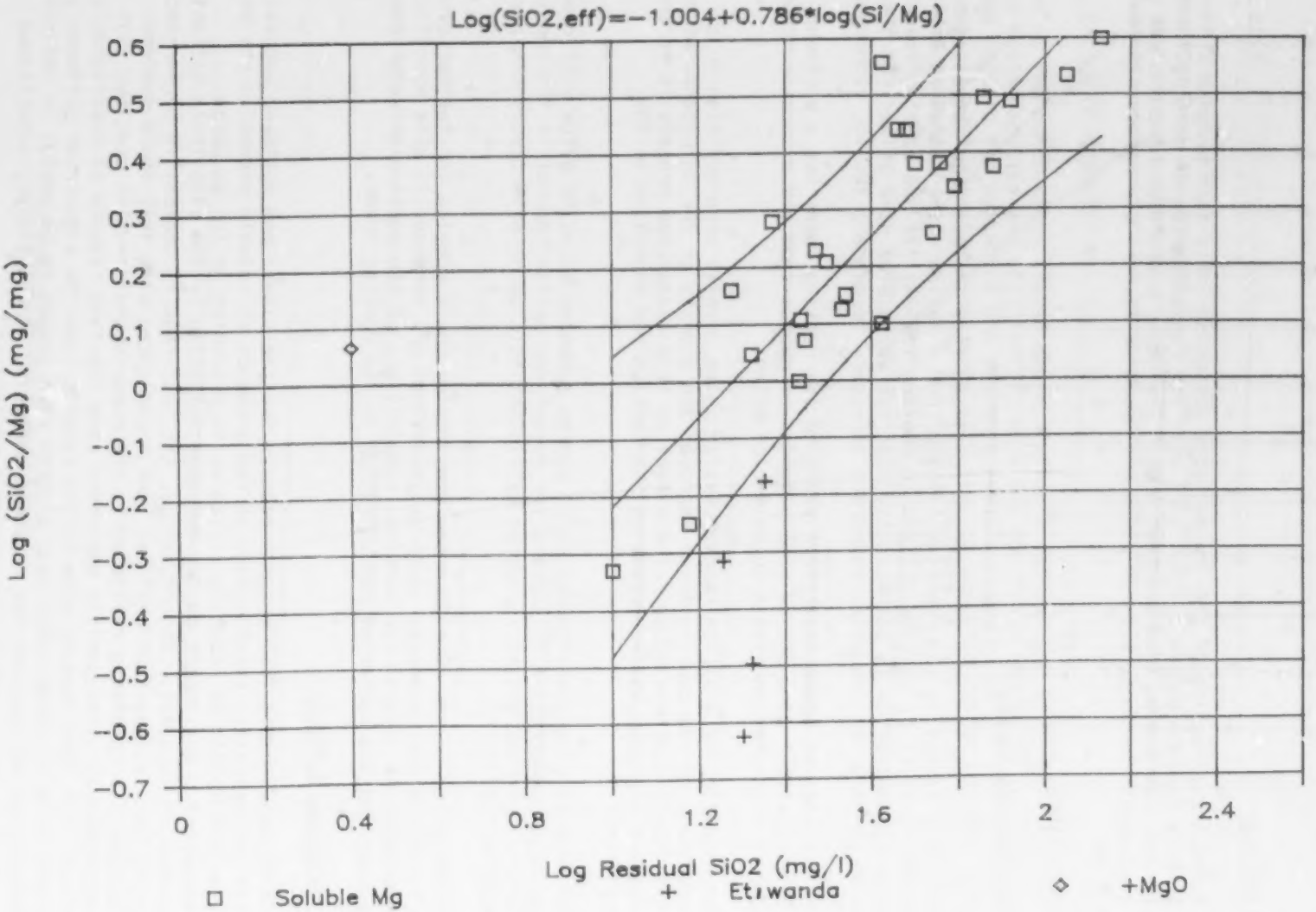
These results reflect a wide range of industrial experience and include a least squares line of best fit, bounded by lines indicating a confidence interval of plus or minus 5%. For further reference, also shown is a data point for the Martin Drake Powerplant which is different because of the addition of solid magnesium. The added magnesium - or more appropriately, its added surface area - is not included with the surface area originating from the precipitation of soluble magnesium. As might be expected, the MgO-point does not fit the trend shown for facilities relying strictly upon soluble magnesium. For soluble magnesium, this plot represents a fairly consistent and coherent pattern of silica removal.

Also, shown are four data points derived from operating results at the Etiwanda field test program. Since no MgO was added, the operating results for a system relying strictly on soluble magnesium are contrary to the broad based industrial experience. The precise reasons for the apparent difference at Etiwanda are unknown. However, the pattern does suggest a performance which is considerably less efficient for the available magnesium and involves something different than what is normally encountered.

A major advantage of pilot systems is that they may point the way to making the full scale system more efficient. For that reason, it is worthwhile to compare performance of similar full scale operating industrial systems to the Etiwanda field test program. For a chemical reactor such as this, comparisons revolve around physical factors that largely determine cost, such as minimizing detention time, chemical feed, effluent concentration, and maximizing factors such as removal and flow rate. The results suggest that the Etiwanda sidestream softener did not compare favorably.

FIGURE 5-2

SIDESTREAM SOFTENER SYSTEM
SILICA REDUCTION VS. AVAILABLE MAGNESIUM



VTFE Unit

It was the intention of the ion exchange softening field test program at Etiwanda to incorporate a small VTFE unit for concentration of cooling tower blowdown wastewater from the test loop. The major evaporator hardware was put in place in the loop, but not piped and operated as part of the overall test system.

CONCLUSIONS

Ion Exchange System

Pilot plant performance at Etiwanda confirmed that ion exchange softening can be highly effective for removing calcium from a saline water source so as to prevent scaling by calcium sulfate in cooling towers. If the cooling tower brine concentration is sufficiently high the waste blow down brine from the tower can provide the sole regenerant necessary to operate the ion exchanger.

The presence of scale inhibitors added to the cooling tower has a profound inhibiting effect also on the precipitation of gypsum that may otherwise accompany Ix regeneration with a sulfate-rich brine.

The operation of a silica removal system using caustic soda addition on a sidestream of the cooling tower improves the quality of the Ix regenerant over what it would otherwise be. The effect of this sidestream process is to increase the sodium while decreasing the calcium and magnesium in the regenerant.

The Ix softening breakthrough point can be extended so as to allow sufficient magnesium penetration to provide for downstream silica removal in the cooling tower sidestream yet still remove sufficient calcium to effectively minimize soda ash addition.

Optimization of the Ix-sidestream combination could provide for further overall process improvements. Such optimizations are necessarily site specific, especially as regards the chemical compositions of the particular water source present and its factor of concentration in the cooling tower.

Sidestream Softener

The Etiwanda field test program results for the sidestream softener suggest that conceptually, caustic soda in the presence of soluble magnesium may be an effective method of reducing silica while producing an ion exchange regenerant. Any impact on ion exchange softening by the sidestream softening appears to be small. Unfortunately, the test results indicate a performance significantly less than what has been experienced with full scale systems. The specific chemistry reflected in the experimental results combined with a non-standard, site-specific apparatus make the test results of questionable value. For that reason, none of the results from the sidestream softener portion of the Etiwanda field test program were judged to be useful in refining the process options for the economic analysis. Conventional, established design data were used in the economic comparison.

VTFE Unit

Because the VTFE was not tested at the Etiwanda Field Test Program, the VTFE unit in Option I was changed to VCE to provide more accurate cost data and to provide a common wastewater evaporator configuration and energy system for the economic comparisons of the five process options.

APPLICATION OF RESULTS

Based on the conclusions stated above, Option I was revised to reflect the updated resin capacity and substitution of a VCE for the original VTFE. The change reduced the number of ion exchange vessels required by 50% and reduced the capital cost directly associated with the ion exchange system by 30%. The economic comparisons in Section 6 are also presented on this basis. Revised process flow and material balances reflecting these changes are shown in Table 5-4. A revised process flow diagram is shown in Figure 5-3.

TABLE 5-4 Ix Softening/Sidestream Silica/VCE/Evaporation Ponds

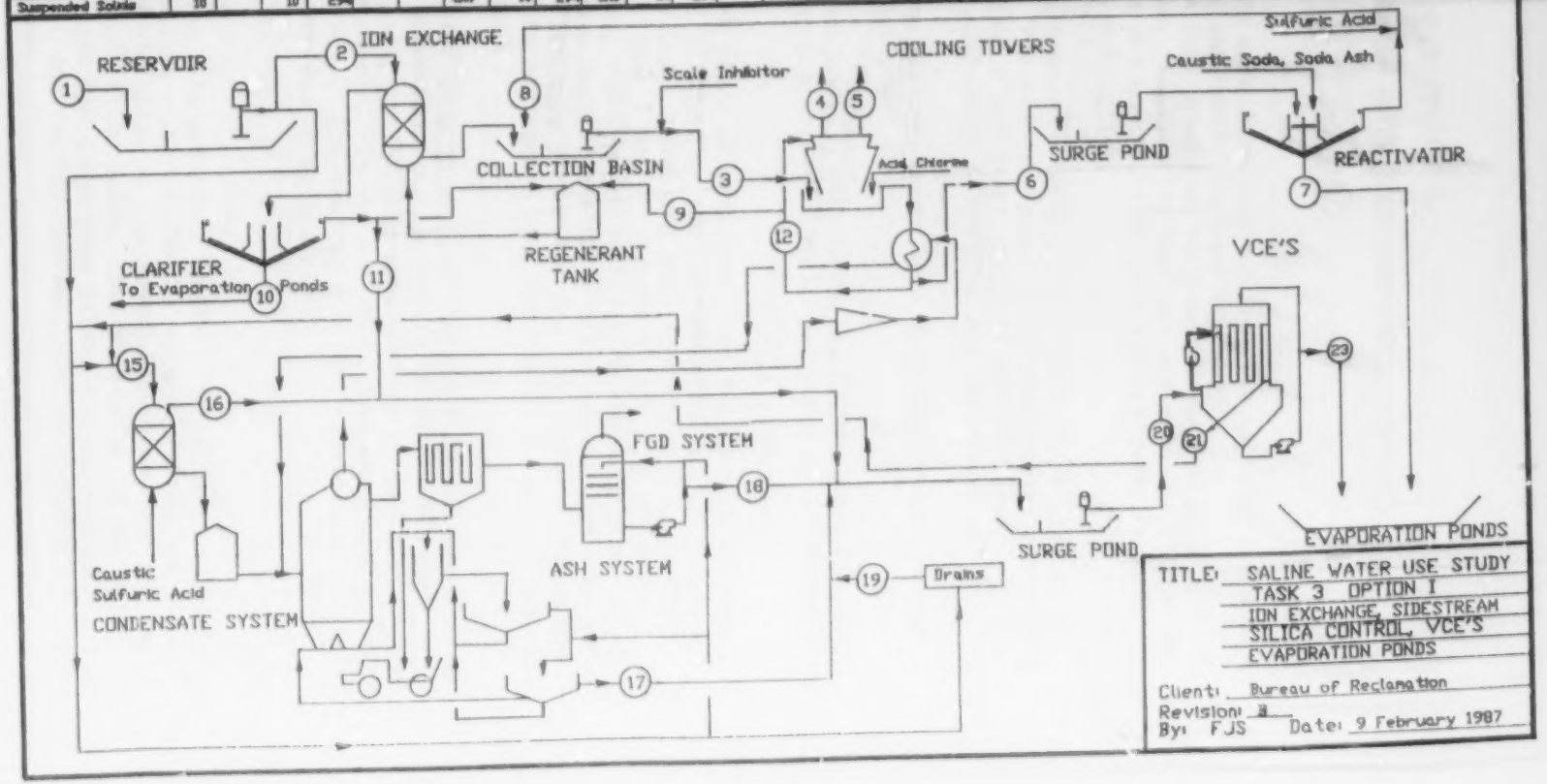
Stream	gpm	Na	Ca	Mg	Cl	SO_4	HCO_3	SiO_2	TDS	SS
Make-Up (Credit Taken for Recovered Water)										
Palo Verde	6557	394	132	46	364	550	322	21	1830	10
Salt,lb/min		21.5	7.2	2.5	19.9	30.1	17.6	1.1	100.1	.5
Ion Exchange										
Feed	6091	394	132	46	364	550	322	21	1830	10
Salt,lb/min		20.0	6.7	2.3	18.5	27.9	16.4	1.1	92.9	.5
Softened Water	6091	518	40	30	364	550	322	21	1845	10
Salt,lb/min		26.3	2.0	1.5	18.5	27.9	16.4	1.1	93.7	.5
Ix Waste Eff	160	13765	600	1028	10542	19956	0	150	46041	20
Salt,lb/min		18.4	.8	1.4	14.1	26.7	0	2.0	61.6	.0
Ix Sludge	20	13765	600	1028	10542	19956	0	150	46041	100000
Salt,lb/min		2.3	.1	.2	1.7	3.3	0	.0	7.6	16.4
Cooling Towers at 29 Cycles of Concentration										
Evaporation	5900									
Drift	20	18305	640	374	10542	26891	80	150	56982	289
Salt,lb/min		3.1	.1	.1	1.8	4.5	.0	.0	9.5	.0
Blowdown	180	18305	640	374	10542	26891	80	150	56982	289
Salt,lb/min		27.5	1.0	.6	15.8	40.4	.1	.2	85.5	.4
SStream Feed	853	18305	640	374	10542	26891	80	150	56982	289
Salt,lb/min		130.2	4.6	2.7	75.0	191.3	.6	.8	405.4	2.1
SStream Out	842	19132	545	1	10542	26985	0	75	57280	10
Salt,lb/min		134.3	3.8	.0	74.0	189.5	0	.5	402.2	.1
SStream Sludge	11	19132	545	1	10542	26985	0	75	57280	100000
Salt,lb/min		1.8	.0	.0	1.0	2.5	0	.0	5.3	7.8
Chemical Feeds										
Caustic (50%)	17.5 lb/min 20.5 lb/kgal Sidestream Feed									
Sulfuric(93%)	15.9 lb/min 18.6 lb/kgal Sidestream Feed									
Soda Ash(98%)	2.0 lb/min 2.3 lb/kgal Sidestream Feed									
Wastewater										
To Treatment										
Ash System	100	400	350	200	400	1921	0	30	3301	150
Regen Waste	50	1970	660	230	1820	2750	0	105	7535	150
FGD Sludge	125	6677	500	200	3640	11000	0	250	22267	10000
Ix Waste Eff	160	13765	600	1028	10542	19956	0	150	46041	20
Miscellaneous	100	394	132	46	364	550	322	21	1830	10
Total to VCE	535	6013	448	422	4320	9445	0	123	20649	20
		26.8	2.0	1.9	19.3	42.2	0	.5	92.2	.1
					7				12	0
Product	485	5						200	170000	16308
Waste	51	52964	400	3717	38046	74673	0	.1	86.2	8.3
Salt,lb/min		26.8	.2	1.9	19.3	37.9				
Final Disposal										
Ix Sludge	20	13765	600	1028	10542	19956	0	150	46041	100000
SStream Sludge	11	19132	545	1	10542	26985	0	75	57280	100000
VCE Waste	51	52964	400	3717	38046	74673	0	200	170000	16308
Total	81	38903	468	2564	27671	54984	0	171	124761	47877

Evaporation Rate at 2 gpm/ac = 40.7 Acres of Disposal Ponds

FIGURE 5-3

REVISED OPTION I PROCESS FLOW DIAGRAM

STREAM	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26	27	28	29
Flow, gpm (lb/min)	6597	6078	6471	20	5900	653	11	642	180	169	20	3900000			300	50	180	123	108	535	495		50						
pH	8.0	S		8.0		S	11.0	8.0	S	8.0	8.0	8.0			5.5	2-14	10	8.0	S	5-6	5-6		5-6						
Sodium mg/l as CO	394	A	518	18305		A	12736	S	A	13745	S	S			5	1940	400	6677	A	6813	5		58264						
Calcium	122	N	40	640		N	545	A	N	680	A	A				640	530	500	N	447			400						
Magnesium	46	C	30	374		C	1	N	C	1028	N	N				230	200	200	C	482			372						
Ammonia		A				A		C	S		C	C							A										
Nitrate	1	S	1			S		A	T		A	S							S										
Chloride	364	I	364	18542		S	10542	L	I	10542	L	I			7	1020	400	3640	I	4320	7		38846						
Bicarbonate	302	T	322	80		T		I	T		I	T							T										
Carbonate		R				R		N	R		N	N							R										
Sulfate	550	E	550	24891		E	2669	E	E	19956	E	E				2750	1901	11000	E	9445			74673						
Phosphate	0.1	A	0.1			A		S	A		S	A							A										
Silica	21	M	21	150		M	75	T	M	150		M				105	30	220	M	122			200						
Dissolved Solids	1830	I	1845	36395		4	37296		4	46041	M				12	7535	3301	22867	I	20649	12		170000						
Suspended Solids	10		10	294			000	10	294	000	20	294				150	150	10000		20			16300						



Section 6

ECONOMIC COMPARISON OF WATER TREATMENT PROCESS OPTIONS

EQUIPMENT COST ESTIMATES

Appraisal level cost estimates were developed for the five selected process options. These costs include equipment, engineering, and installation/construction. A cost breakdown is provided for the major components of each process option. Engineering cost was estimated at 20% of the equipment cost. Installation and construction cost was estimated as 1.75 times the equipment cost for each option.

The capital cost estimates were made on an appraisal level basis with an accuracy of plus or minus 25%. For standard power plant components, costs were developed using data from recent power plant projects and from Stearns Catalytic's power plant cost estimating procedures and data base.

For the ion exchange softening options, costs were developed in conjunction with ion exchange resin and equipment suppliers.

VCE costs were developed in conjunction with Resources Conservation Company, based on the cost of VCE brine concentrators recently installed at Mohave Station.

Condenser materials costs were based on estimates from tubing material suppliers for the range of circulating water chemistries expected for the five options. Inhibited admiralty brass tubes are used in Options III, IV and V. Titanium tubes are used in Options I and II because of the high chloride levels. In the cost estimates, the upgrade to titanium tubes is shown as an incremental cost above the base cost of admiralty brass tubes. The cost for inhibited admiralty is $3.23 per square foot for one inch diameter tubes and 20 gauge thickness versus $7.26 per square foot for titanium tubes of one inch diameter and 22 gauge thickness.

Evaporation pond costs are based on a total pond cost of $125,000 per acre for a single-lined pond with an underdrain and monitoring system. This cost is based on an update of recent evaporation pond construction at San Juan Station and at other analogous power plants. In the estimate, pond materials and land are estimated at $42,400 per acre. Engineering and construction are estimated at the same factors as used for the other equipment.

Option I

The major equipment for Option I is presented below. The battery limits for the ion exchange softening system start at the raw water inlet to the ion exchange softening system, the discharge to the cooling tower basins, the wastewater surge pond and the evaporation ponds. The boundaries of the wastewater treatment system include the surge pond inlet, the make-up points to the condensate system, the ash and the FGD systems, and the evaporation ponds.

Titanium is substituted for the inhibited admiralty tubes due to the high levels of chlorides predicted in the circulating water system. The estimated additional tubing cost for the two condensers, at approximately 360,000 square feet, is $1,500,000.

1. Ion Exchange Softening

   (6) 10 ft. diameter by 15 ft. high lined steel vessels with all necessary piping, instrumentation and controls, electrical, regenerant and rinse tanks and instrumentation
   (1) 2.5 mgal synthetically lined collection basin with (3) 6,000 gpm pumps, electrical, piping and instrumentation

   (1) 1 mgal, synthetically lined softener surge pond with (3) 500 gpm pumps, electrical and instrumentation

   (2) 600 gpm contact clarifiers with caustic and soda ash storage and feed systems, acid storage and feed system, piping, electrical and instrumentation

2. Wastewater Treatment

   (1) 5 mgal, synthetically lined surge pond with (3) 300 gpm pumps

   (2) 350 gpm VCE's with feed tank, deaeration, instrumentation, electrical, control room, chemical feed equipment, and distillate tank

   41 acres of synthetically lined evaporation ponds - 15 to 20 feet deep

Option II

The major equipment for Option II is presented below. The boundaries for the sidestream softening system include the inlet to the softener surge pond, the discharge from the neutralization pond pumps, the inlet to the wastewater surge pond and the inlet to the evaporation ponds. The wastewater treatment boundaries includes the inlet to the surge pond, the make-up to the condensate system and the evaporation ponds.

Titanium is substituted for the inhibited admiralty tubes due to the high levels of chlorides predicted in the circulating water system. The estimated additional tubing cost for the two condensers, at approximately 360,000 square feet, is $1,500,000.

1. Sidestream Softening

   (1) 1 mgal, synthetically lined softener surge pond with (3) 750 gpm pumps, electrical and piping

   (2) 1000 gpm softeners with caustic and soda ash storage and feed systems, acid storage, coagulant, polymer storage and feed systems, piping, electrical and instrumentation

   (1) 1 mgal, synthetically lined neutralization pond with (3) 500 gpm pumps, electrical and piping

2. Wastewater Treatment

   (1) 5 mgal, synthetically lined surge pond with (3) 300 gpm pumps

   (2) 350 gpm brine concentrators with acid storage and feed system, polymer blending and feed, deaeration, electrical, pumps, piping, instruments, control room, and product storage.

   41 acres synthetically lined evaporation ponds 15 to 20 feet deep

Option III

The major equipment for Option III is presented below. The boundaries for Option III includes only the evaporation ponds.

1. Wastewater Treatment

   845 acres of synthetically lined evaporation ponds 15 to 20 feet deep

Option IV

The major equipment for Option IV is presented below. The boundaries for the process include the inlet to the wastewater surge pond, the make-up inlets to the condensate make-up, ash, FGD and cooling tower systems and the Evaporation Ponds.

1. Wastewater Treatment

(1) 5 mgal synthetically lined surge pond with (3) 900 gpm pumps

(4) 500 gpm VCE'S with acid storage and feed system, polymer blending and feed, deaeration, electrical, pumps, piping, instruments, control room, and product storage.

31 acres of synthetically lined evaporation ponds 15 to 20 feet deep

Option V

The major equipment for Option V is presented below. The boundaries for the softening system include the raw water inlet to the softeners, the inlet to the cooling tower basins and the inlet to the Evaporation Ponds.

1. Make-Up Softening

(2) 5,000 gpm capacity softeners with lime storage and slakers, soda ash, acid coagulant storage and feed systems, thickener, pumps, piping, electrical and instrumentation

(1) 2.5 mgal concrete lined collection basin with (3) 6,000 gpm pumps, electrical, piping and instrumentation

2. Wastewater Treatment

(1) 5 mgal synthetically lined surge pond with (3) 500 gpm pumps

(2) 500 gpm VCE's with acid storage and feed system, polymer blending and feed, deaeration, electrical, pumps, piping, instruments, control room, and product storage.

24 acres of synthetically lined evaporation ponds 15 to 20 feet deep

Equipment Cost Summary for Options I - V

The total capital costs by cost element are summarized in Table 6-1 for the five options. Option III has the highest estimated capital cost of $105.7 million. The remaining options are significantly less with Option II the least at $22.4 million. Options I, IV and V range from $26.0 million to $33.3 million.

TABLE 6-1

SUMMARY OF TOTAL CAPITAL COSTS
($1,000)

Cost Element	Options I	II	III	IV	V
Equipment					
Ion Exchange	1,190				
Make-Up Softening					3,800
Sidestream	1,050	1,400			
Surge Ponds	850	450		350	750
VCE's	3,500	3,500		9,500	5,750
Evaporation Ponds	1,725	1,725	35,832	1,309	996
Total Equipment	8,315	7,075	35,832	11,159	11,296
Installation Cost	14,551	12,382	62,706	19,529	19,769
Engineering Cost	1,663	1,415	7,166	2,232	2,259
Condenser Tubing	1,500	1,500			
Total Capital	26,029	22,372	105,704	32,920	33,324

ECONOMIC ANALYSIS OF PROCESS OPTIONS

Levelized cost comparisons of the five selected process options were developed using integrated microcomputer spreadsheet software. Economic factors for the analysis were provided by SCE and reviewed by PG&E. The economic analysis categories included capital cost, operating and maintenance cost, energy and demand cost and raw material cost.

Economic Assumptions and Model Description

The economic model is cross-linked with the process flow and material balance spreadsheets, the raw water chemistry data base, the equipment costs and the economic factors. The assumptions for chemical cost in 1985 dollars are presented in Table 6-2. The chemical costs are based on trucked freight to the reference plant.

TABLE 6-2

CHEMICAL COSTS

Item	1985 Cost in Dollars
Lime (85% CaO)	80 per ton
Soda Ash (98% Na2CO3)	120 per ton
Caustic Soda (50% NaOH)	300 per ton
Sulfuric Acid (93% H2SO4)	40 per ton
Chlorine	300 per ton
Polymers, Scale Inhibitors	1.25 per pound

The chemical costs for each option are based on the usage in pounds per thousand gallons as determined in the process flow and material balances. Operating labor costs for a given option are developed by assuming the number of operators and supervisors required for a given system and multiplying by the labor cost for a manyeer. The repair materials cost is assumed to be 0.75% per cent of the total capital cost.

Operating labor cost is assumed at $50,000 per manyear. Maintenance labor cost is assumed as 40% of the repair materials estimate.

Economic factors for the study were provided by SCE and reviewed by PG&E. Factors for levelizing capital, operating and maintenance costs, plus energy cost and demand penalty are presented in Table 6-3. These factors are applied to the total capital costs, the operating and maintenance yearly costs, power usage per thousand gallons treated, or handled, and the connected power demand for each option.

TABLE 6-3

ECONOMIC FACTORS
(January, 1985 dollars)

Levelized Capital Charge Rate:	24%
O & M Levelizing Factor:	1.65
Levelized Energy Usage Cost:	$.047/kwh
Energy Demand Cost:	$145 per kw-yr
Plant Life:	30 years

The economic model is divided into six sections. The first section summarizes the flows to be treated or handled by the different systems within each option, plus the evaporation pond acreage and the quantity of recovered water. The second section tabulates the chemical usage and yearly costs for each option. The operating and maintenance labor and repair material costs are summarized in the third section as 1985 dollars per year. The fourth section is an equipment, installation and engineering cost summary in 1985 dollars. Section five lists the power consumption, energy cost, demand cost and total power costs for each option. The final section presents the levelized cost elements for capital, chemicals, operating and maintenance labor and repair,

the total power costs and the total levelized costs.

Economic Comparisons

The results of the economic comparisons of the five selected process options are presented in Table 6-4.

TABLE 6-4

ECONOMIC COMPARISON OF PROCESS OPTIONS

Section I -- Water Treated (gpm)

Option	I	II	III	IV	V
With Ion Exchange	6091	0	0	0	0
With Sidestream Treatment	853	1368	0	0	0
With Make-Up Softening	0	0	0	0	6666
With Brine Concentrators	535	555	0	1615	868
Disposal Pond Acres	41	41	845	31	24

Section II -- Chemical Usage and Costs ($1000 per year)

85% Lime					
lb/kgal	0	4.6	0	0	3.8
Cost	0	134	0	0	543
50% Caustic					
lb/kgal	20.5	0	0	0	0
Cost	1388	0	0	0	0
93% Sulfuric Acid					
lb/kgal	18.6	11.7	2.2	2.2	2.5
Cost	168	170	194	154	176
98% Soda Ash					
lb/kgal	2.3	18.8	0	0	3.2
Cost	73	956	0	0	781
Chlorine					
lb/kgal	.03	.03	.03	.03	.03
Cost	55	50	66	53	53
Polymers					
lb/kgal	.03	.03	.03	.03	.03
Cost	138	143	165	133	132
Total Chemical Cost	1822	1453	425	340	1685

Section III -- Operating Labor, Maintenance and Materials ($1,000 per year)

Option	I	II	III	IV	V
Operating Labor	600	500	200	300	500
Maintenance Labor	75	67	317	99	100
Repair Materials	195	168	793	247	250
Total Labor & Repair	870	735	1310	646	850

Section IV -- Equipment, Engineering and Installation Cost ($1,000)

Equipment Cost					
Ion Exchange	1190				3800
Make-Up Softening					
Sidestream Treatment	1050	1400			
Collection/Surge Ponds	850	450		350	750
VCE'S	3500	3500		9500	5750
Evaporation Ponds	1725	1725	35832	1309	996
Total Equipment Cost	8315	7075	35832	11159	11296
Engineering Costs	1663	1415	7166	2232	2259
Installation Cost	14551	12382	62706	19529	19769
Condenser Materials	1500	1500			
Total Capital Costs	26029	22372	105704	32920	33324

Section V -- Power Consumption and Demand

Power Usage					
Kwh per kgal @ Towers	5	5	0	0	5
Kwh per Kgal @ Wastewater	90	90	5	90	90
Cost ($1000/year)	2225	1415	211	3621	2776
Power Demand					
Kw Demand	2766	4259	634	10903	8359
Cost ($1000/year)	928	618	92	1581	1212

Section VI -- Levelized Annual Cost Summary ($1000)

Levelized Capital	6247	5369	25369	7901	7998
Levelized Chemicals	3006	2397	701	561	2780
Levelized Labor & Repair	1436	1213	2162	1066	1403
Power	3153	2032	302	5202	3988
Total Levelized Cost	13842	11011	28534	14730	16169

Option II is the most cost-effective with a levelized annual cost of $11 million. Option I, the Etiwanda field test unit configuration, is second at $13.8 million. Options IV at $14.7 and Option V at $16.2, which are the

proven technology options, are close in third and fourth positions. Option III, with evaporation ponds only for wastewater treatment, is the most costly at $28.5 million.

Option II has an indicated 25 - 30% savings potential over Option IV and V, due to its lower capital and power costs. Option I has an indicated savings of 6 - 15% over Option's IV and V. The lower capital cost is offset by the increased chemical cost for the silica control system and the power cost for wastewater treatment.

SENSITIVITY ANALYSIS OF PROCESS OPTIONS

The major cost components of each option were varied from 75 per cent to 150 per cent of base case values to determine areas of economic sensitivity. Components selected for sensitivity analysis included chemical costs, labor and repair cost, total capital cost and power cost. The levelized cost results of the evaluation are presented in Table 6-5.

TABLE 6-5

SENSITIVITY ANALYSIS RESULTS ($1000)

(Levelized Cost in $1000)

Option	I	II	III	IV	V
Vary Chemicals					
75% of Base	13,190	10,412	28,359	14,590	15,474
Base Value	13,842	11,011	28,534	14,730	16,169
150% of Base	15,695	12,210	28,885	15,011	17,559
Vary Labor & Maintenance					
75% of Base	13,566	10,708	27,994	14,464	15,818
Base Value	13,842	11,011	28,534	14,730	16,169
150% of Base	14,634	11,618	29,625	15,263	16,870
Vary Power					
75% of Base	13,134	10,503	28,459	13,430	15,172
Base Value	13,842	11,011	28,534	14,730	16,169
150% of Base	15,499	12,027	28,685	17,331	18,163
Vary Capital					
75% of Base	12,268	9,669	22,192	12,755	14,170
Base Value	13,842	11,011	28,534	14,730	16,169
150% of Base	16,954	13,696	41,219	18,680	20,168

The results of the sensitivity analysis show that varying power cost, capital or labor and maintenance cost does not change the order of the base case ranking.

Option I slips to third place if the chemical costs are increased 150% over

the base case, indicating that the chemical requirements of the sidestream silica control system can affect the cost competitiveness of Option I significantly. Options I, II and V are sensitive to varying chemical costs and/or usage since all three rely on chemical processes to modify the circulating water chemistry. Options III and IV are relatively insensitive because the configurations do not rely on chemical processes.

Those options using VCE's for wastewater treatment are sensitive to power usage and cost. Option I is more sensitive than Option II because of the double handling required for the cooling tower make-up and cooling tower blowdown at the ion exchange system.

Section 7

CONCLUSIONS AND RECOMMENDATIONS

CONCLUSIONS

The objective of the Task 3 study was to compare the cost of ion exchange softening technology with alternative water treatment process options in the context of a complete coal-fired power plant supplied by a saline water source from the Colorado River Basin. Supplemental to the study was the Etiwanda Field Test Program which provided design and operating data for ion exchange softening technology, using cooling tower blowdown for regeneration of the ion exchange resin, and operating on a synthesized quality similar to that assumed in the study.

The water quality selected for the study was the Palo Verde Irrigation Outfall Drain, which has sodium to calcium ratios favorable to ion exchange softening, but also has silica levels which require sidestream silica control in the plant circulating water system. Palo Verde Irrigation Outfall Drain water was selected because it had a sound water quality data base and could actually be used for a power plant sited in the Colorado River corridor in the California/Nevada/Arizona border area.

The Etiwanda Field Test Program originally intended to operate with more than one water quality and also intended to evaluate VTFE technology through the pilot plant tests. As the program turned out, only the simulated Palo Verde Irrigation drainage water was tested and the VTFE was not incorporated into the test loop. The field test program was successful in its principal objective of confirming the technical viability of ion exchange softening using cooling tower blowdown in either an upflow fluidized bed mode or rapid downflow fixed bed more for resin regeneration. The test program also concluded that the ion exchange softener capacity, using cooling tower blowdown for regeneration, was higher than assumed in the initial design analysis for the study. Thus, the field test program provided valuable inputs to the final economic comparisons of water treatment process options.

Levelized cost comparisons of ion exchange softening technology with four alternative process options showed that sidestream lime/soda softening was approximately 26% more cost-effective for the Palo Verde Irrigation Outfall Drain water quality. Ion exchange softening was second in overall cost-effectiveness. The economic and sensitivity analysis results show that the cost of silica reduction in the ion exchange softening process option offsets the cost savings of using cooling tower blowdown for regeneration. The economic analysis results point out that the Palo Verde water quality is not optimum for ion exchange softening technology, but that ion exchange softening is likely to be cost-competitive for other waters of high hardness and lower silica concentrations.

The study also pointed out the importance of the role of wastewater evaporators to reduce evaporation pond area and to provide a high quality makeup water source for condensate systems, FGD systems and other plant uses.

VTFE technology was not included in the final economic comparisons because the VTFE system did not end up being tested at Etiwanda. In the initial design and economic analysis phase of the study, appraisal-level design and cost estimating was performed for the application of VTFE systems using condenser exhaust steam. Cost assumptions for steam line design are included in Appendix 8-2. The preliminary cost data showed that wastewater evaporators driven by waste heat steam can offer a substantial cost savings if the practical issues of steam availability, steam line length and actual steam cost are clarified. In a new plant case, these factors can be incorporated into the design. In a retrofit case, they will be highly site-specific.

In the development of process design options, operation of the plant circulating system at high cycles of concentration was generally cost-effective. At these high concentrations, the corrosion impacts of constituents such as chloride throughout the plant cooling water system could be substantial. In the study, higher quality materials of construction were provided for condenser tubing and cooling tower materials, however, complete maintenance cost impacts would require a more detailed materials analysis.

RECOMMENDATIONS

Based on the results of the technical and economic comparisons of water treatment process options and the results of the Etiwanda Field Test Program, the following recommendations are made:

1. Ion exchange softening technology should be tested on a demonstration plant scale if site-specific power plant design criteria and water supply quality show an economic advantage. The test program would be enhanced by the involvement of both ion exchange resin manufacturers and ion exchange equipment suppliers.

2. Sidestream lime/soda softening technologies should be demonstrated on high salinity waters to develop accurate operating cost data. Lime/soda softening technologies showed a strong economic position in the Task 1, Task 2 and Task 3 studies. Sidestream lime/soda softening has been particularly strong, offering capital, chemical and power cost advantages in many cases. Sidestream lime/soda and caustic/soda softening has been demonstrated on moderate salinity waters. Field testing and model confirmation on cooling towers operating on high salinity makeup water sources would provide valuable operating experiences and cost data.

3. Waste heat evaporators for concentration and reuse of powerplant wastewaters should be evaluated on a design-level basis to determine accurate economic comparisons with vapor compression evaporators. Adequate data have been developed for the process design of VCE and VTFE evaporators, however, a comprehensive assessment of the practical in-plant issues of waste heat steam supply system design and cost would benefit from additional engineering studies.

4. Power plant cooling water system materials of construction should be evaluated for application in high chloride and high total dissolved solids operating environments. Evaluations should consider not only the

condenser and cooling tower equipment, but all auxiliary cooling systems and the effects of highly concentrated cooling tower drift on plant equipment maintenance and regulatory requirements.

Section 8

APPENDIX

Appendix		Page
8-1	Coal Quality Analysis	8-2
8-2	Capital Cost Estimates for Ix, Condenser Tubing and VTFE Steam Piping	8-3

APPENDIX 8-1

REFERENCE POWER PLANT COAL QUALITY

ULTIMATE ANALYSIS

	As Received	Dry Basis
%Moisture	3.27	xxxxx
%Carbon	72.96	75.43
%Hydrogen	5.00	5.17
%Nitrogen	1.35	1.40
%Chlorine	0.01	0.01
%Sulfur	0.49	0.51
%Ash	7.04	7.28
%Oxygen (diff.)	9.88	10.20
Total	100.00	100.00

MINERAL ANALYSIS OF ASH	%Wt. Ignited Basis
Phosphorous pentoxide	0.99
Silica as SiO2	50.86
Ferric oxide	4.21
Alumina	16.10
Titania	0.80
Lime	10.88
Magnesia	3.71
Sulfur trioxide	9.80
Potassium oxide	0.57
Sodium oxide	0.93
Undetermined	1.15
Total	100.00

APPENDIX 8-2

Capital Cost Estimates for ion exchange softening, condenser tubing and VTFE steam piping.

Stearns Catalytic

STEARNS CATALYTIC CORPORATION

November 8, 1985

Mr. Jack K. Laughlin
Consulting Engineer
P.O. Box 626
Borrego Springs, California 92004

Subject: Capital Cost Estimates
BuRec Task 3
Project No. 27835

Dear Jack:

Enclosed are capital cost estimates for items you have requested. The capital costs for the Ion Exchange Systems are as requested in Rick Schroeder's letter of 10/11/85 and as stated herein. The following is a data summary. If required, support calculations are available.

Ion Exchange Softeners

Equipment estimates for the ion exchange softener systems were provided by two major equipment suppliers. Both suppliers have technical research departments and extensive experience in a broad range of ion exchange applications.

Initially the vendors were presented the performance data contained in Rick Schroeder's letter of 10/11/85. They were asked to provide price estimates for systems that would meet the requirements of the two cases. Shortly after being presented with the data, the vendors expressed considerable doubt the anticipated performance could be achieved on a commercial basis. Both vendors were reluctant to expend any additional effort unless supportive data was provided. In particular, the anticipated capacity of the resin under the stated operating conditions was needed.

Next, Rohm & Haas was contacted and asked to estimate the capacity of Amberlite IR-120 under the specified operating conditions. On the basis of the information presented, Rohm & Haas also questioned the feasibility of the approach and declined to make an estimate of resin capacity.

At this point, it was decided SCC would use resin capacity data developed by John Kaakinen.. SCC would then calculate resin volumes and vessel sizes based on this data. The sizing would be for a conventional sodium cycle cation exchange system using utility type design factors. After the initial sizing was completed, it was reviewed with John and mutual agreement was reached. The design factors used were at the high end of

4500 Cherry Creek Drive, P.O. Box 5888, Denver, Colorado 80217, (303) 758-1122, TLX 45540, TWX 910-931-0453

the ranges, but still within established operating practices. The vendors were then asked to provide estimates for equipment as sized by SCC. Both vendors were agreeable to do so, but asked that their names not be used on the estimates. They were concerned that providing estimates for this equipment in their name might imply concurrence with the technical viability of the scheme.

The estimates for Case 1 and Case 2 are detailed on the attached Estimate Summary sheets and are summarized as follows:

	Case 1	Case 2
Equipment	$1,980,250	$ 731,225
Installation	217,875	115,315
Engineering	240,035	240,035
Total	$2,438,160	$1,086,575

The prices include the specified equipment installed in a prefabricated building. The estimate assumes no other equipment other than the basic water treating package. From our experience on other water treating packages, Stearns feels this estimate is substantially lower than expected. This is probably due in part to the vendors doubting the viability of the scheme and thereby giving minimum attention to the details of the estimate. The remaining deficiency is most likely a result of the abbreviated scope of work. If this estimate were part of a conventional preliminary engineering effort, a substantial amount of additional work would be identified. It is Stearns' opinion these estimates could easily double if all the systems were fully designed and all associated areas were completely defined.

No operating or maintenance costs are included. Resin replacement cost can be calculated at $50/cubic foot.

The following is a summary of the operating parameters for the ion exchange systems as established by John Kaakinen (BuRec), Rick Schroeder, Bob Hamilton (SCC), and the equipment suppliers.

Ion Exchange Cycle

Service	283 minutes (4.72 hours)
Regeneration	73 minutes (1.22 hours)
Total	356 minutes (5.94 hours)

Resin Capacity and Type

85 Bed Volumes of Water
Amberlite IR-120

Vessels	Case 1	Case 2
Total Flow Rate, GPM	10,000	3,500
Number of Vessels	12	4
Total Resin Volume, Cu. Ft.	4,451	1,557
Resin per Vessel, Cu. Ft.	371	389
Vessel Diameter, Ft.	10	10
GPM/Sq. Ft.	10.6	11.1
GPM/Cu. Ft.	2.25	2.25
Bed Depth, Ft.	4.7	5.0
Minimum Bed Expansion, %	50	50
Overall Height, Ft.	15	15
Filtration	None	None

Condenser Tube Materials

Price estimates for three different condenser tube materials are provided - inhibited admiralty, 90-10 copper nickel, and titanium. The square footage required for each tube material was calculated on the basis of the physical properties of each material. Factors such as heat transfer rate, fouling factor, allowable velocity and strength were part of the sizing basis. Prices are FOB Denver.

Inhibited Admiralty - ASTM B111 Seamless

177,000 Square Feet
1" OD x 18 BWG (0.049")
28,200 Tubes
$84.45/100 LF; $3.23/Square Foot
$571,710 Total

90-10 Copper-Nickel - ASTM B111 Welded

186,000 Square Feet
1" OD x 20 BWG (0.035")
25,600 Tubes
$115.22/100 LF; $4.40/Square Foot
$818,400 Total

Titanium - ASTM B338 Welded

183,000 Square Feet
1" OD x 22 BWG (0.028")
20,000 Tubes
$190.00/100 LF; $7.26/Square Foot
$1,328,580 Total

Steam and Condensate Lines

Cost Summary Sheets are provided for 2,000 foot lengths of steam and condensate lines. These are installed costs with the lines routed above grade on sleepers. It is assumed largely straight runs will exist. No valves, expansion joints or condensate collection systems are included. Insulation (but no heat tracing) is included on the steam lines but not the condensate line. Steam line conditions were calculated on the basis of steam at 1.473 psia and 115°F being routed through 2,000 feet of pipe. A summary of operating conditions is as follows:

Pipe Size ID, In.	48"	60"
Pressure Drop, PSI/100 Ft.	0.017	0.0056
Velocity, FPM	23,030	14,740
End of Pipe Temperature, °F	106	112

It should be noted that at certain times of the year, the ambient air temperature may be hotter than the steam supplied to the VTFE.

Start-up and shutdown of this pipeline is expected to be difficult considering the low temperature of the steam and resultant large quantities of condensate which must be removed from the line. Also, the integrity of the power cycle must be maintained. Sufficient valving, instrumentation, etc. must be incorporated to assure this.

If you need clarification of the information presented or additional information, please contact me.

Sincerely,

STEARNS CATALYTIC CORPORATION

R L Hamilton

R. L. Hamilton

RLH/md
Enclosures

cc: F. J. Schroeder
A. W. McKenzie
W. E. Ward
J. J. Donovan

1499d(0048d)

Steams Catalytic

ESTIMATE SUMMARY

CLIENT	BuREC - Task 3 - Case 1	JOB NO.	27835
LOCATION	Blythe, California	DATE	11-1-85
PROJECT	Water Treating Building	BY	A. Cole
REV. NO.	REV. DATE	BY	

ACT	DESCRIPTION	CRAFT HOURS	LABOR	MATERIAL	OTHER	TOTAL
A	EARTHWORK	162	2 830			2 830
B	CONCRETE	492	8 170	6 830		15 000
C	BUILDINGS & STRUCTURES				103000	103 000
D	PROCESS EQUIPMENT	480	9 310	1460 500		1469 810
E	PIPING	284	5 570	20 000		25 570
F	ELECTRICAL	215	4 360	4 070		8 430
G	PAINTING					
L	PLANT FACILITIES					
N	INSTRUMENTS & CONTROLS	36	740	700		1 440
P	INSULATION					
T	DEMOLITION					
	DIRECT FIELD COST	1 669	30 980	1492 100	103000	1626 080
H	FIELD STAFF		6 200			6 200
H8	LEGALITIES		7 100			7 100
K	CONSTRUCTION SUPPLIES			1 700		1 700
M	STARTUP		1 550			1 550
S	TEMPORARY FACILITIES		3 100	700		3 800
V	CRAFT BENEFITS		7 750			7 750
V8	CONSTRUCTION CAMP					
W	CONSTRUCTION EQUIP.				4 000	4 000
	INDIRECT FIELD COST		25 700	2 400	4 000	32 100
	TOTAL FIELD COST		56 680	1494 500	107 000	1658 180
J	HOME OFFICE (4378 Hours)					194 030
	TOTAL FIELD & HOME OFFICE					1852 210
O	TAXES					89 700
R	PREMIUM PAY					620
Y3	CONTINGENCY					485 630
Y4	ESCALATION					
	SUB TOTAL					2428 160
Y2	FEES					10 000
	TOTAL					2438 160

THIS ESTIMATE IS BASED UPON CONTRACTOR'S EXPERIENCE, INCLUDING STATED RANGES OF ACCURACY. THE PROBABILITY OF PROJECT COST BEING WITHIN THE STATED ACCURACY RANGE IS ___%; CONVERSELY THE PROBABILITY OF PROJECT COST OUTSIDE THE STATED ACCURACY RANGES IS___%. THEREFORE, CONTRACTOR DOES NOT ASSUME THE RISK OF COSTS VARYING FROM THOSE ESTIMATED.

FORM 32-11FSC
(REPLACES FORM 32-11F)

Steams Catalytic

ESTIMATE SUMMARY

CLIENT	BuREC - Task 3 - Case 2	JOB NO.	27835
LOCATION	Blythe, California	DATE	11-1-85
PROJECT	Water Treating Building	BY	A. Cole
REV. NO.	REV. DATE	BY	

ACT	DESCRIPTION	CRAFT HOURS	LABOR	MATERIAL	OTHER	TOTAL
A	EARTHWORK	76	1 330			1 330
B	CONCRETE	261	4 330	3 625		7 955
C	BUILDINGS & STRUCTURES				58600	58 600
D	PROCESS EQUIPMENT	240	4 650	536 500		541 150
E	PIPING	104	2 040	8 000		10 040
F	ELECTRICAL	76	1 550	1 505		3 055
G	PAINTING					
L	PLANT FACILITIES					
N	INSTRUMENTS & CONTROLS	16	325	300		625
P	INSULATION					
T	DEMOLITION					
	DIRECT FIELD COST	773	14 225	549 930	58600	622 755
H	FIELD STAFF		2 900			2 900
H8	LEGALITIES		3 300			3 300
K	CONSTRUCTION SUPPLIES			800		800
M	STARTUP		700			700
S	TEMPORARY FACILITIES		1 500	200		1 700
V	CRAFT BENEFITS		3 600			3 600
V8	CONSTRUCTION CAMP					
W	CONSTRUCTION EQUIP				2000	2 000
	INDIRECT FIELD COST		12 000	1 000	2000	15 000
	TOTAL FIELD COST		26 225	550 930	60600	637 755
J	HOME OFFICE (4378 hours)					194 030
	TOTAL FIELD & HOME OFFICE					831 785
Q	TAXES					33 050
R	PREMIUM PAY					425
Y3	CONTINGENCY					216 315
Y4	ESCALATION					
	SUB TOTAL					1081 575
Y2	FEES					5 000
	TOTAL					1086 575

THIS ESTIMATE IS BASED UPON CONTRACTOR'S EXPERIENCE, INCLUDING STATED RANGES OF ACCURACY. THE PROBABILITY OF PROJECT COST BEING WITHIN THE STATED ACCURACY RANGE IS ___%. CONVERSELY THE PROBABILITY OF PROJECT COST OUTSIDE THE STATED ACCURACY RANGES IS___%. THEREFORE, CONTRACTOR DOES NOT ASSUME THE RISK OF COSTS VARYING FROM THOSE ESTIMATED.

FORM 82-11FSC
(REPLACES FORM 82-11F)

Stearns Catalytic

ESTIMATE SUMMARY

CLIENT	BuREC - Task 3	JOB NO.	27835
LOCATION	Blythe, California	DATE	11-4-85
PROJECT	2000 Feet of 60" Pipe	BY	A. Cole
REV. NO.	REV. DATE	BY	

ACT	DESCRIPTION	CRAFT HOURS	LABOR	MATERIAL	OTHER	TOTAL
A	EARTHWORK					
B	CONCRETE					
C	BUILDINGS & STRUCTURES					
D	PROCESS EQUIPMENT					
E	PIPING	16 932	332 200	274 820		607 020
F	ELECTRICAL					
G	PAINTING					
L	PLANT FACILITIES					
N	INSTRUMENTS & CONTROLS					
P	INSULATION				89 930	89 930
T	DEMOLITION					
	DIRECT FIELD COST	16 932	332 200	274 820	89 930	696 950
H	FIELD STAFF					
H8	LEGALITIES					
K	CONSTRUCTION SUPPLIES					
M	STARTUP					
S	TEMPORARY FACILITIES					
V	CRAFT BENEFITS					
V6	CONSTRUCTION CAMP.					
W	CONSTRUCTION EQUIP.					
	INDIRECT FIELD COST					305 890
	TOTAL FIELD COST					1 002 840
J	HOME OFFICE					69 695
	TOTAL FIELD & HOME OFFICE					1 002 840
Q	TAXES					18 440
R	PREMIUM PAY					6 640
Y3	CONTINGENCY					164 640
Y4	ESCALATION					
	SUB TOTAL					1 262 255
Y2	FEES					55 615
	TOTAL					1 317 870 *
	* $659.00 per linear foot					

THIS ESTIMATE IS BASED UPON CONTRACTOR'S EXPERIENCE, INCLUDING STATED RANGES OF ACCURACY. THE PROBABILITY OF PROJECT COST BEING WITHIN THE STATED ACCURACY RANGE IS ___%; CONVERSELY THE PROBABILITY OF PROJECT COST OUTSIDE THE STATED ACCURACY RANGES IS___%. THEREFORE, CONTRACTOR DOES NOT ASSUME THE RISK OF COSTS VARYING FROM THOSE ESTIMATED.

FORM 82-11FSC
(REPLACES FORM 82-11F)

Stearns Catalytic

ESTIMATE SUMMARY

CLIENT	BuREC	JOB NO.	27835
LOCATION	Blythe, California	DATE	11-4-85
PROJECT	2000 Feet of 48" Pipe	BY	A. Cole
REV. NO.	REV. DATE	BY	

ACT	DESCRIPTION	CRAFT HOURS	LABOR	MATERIAL	OTHER	TOTAL
A	EARTHWORK					
B	CONCRETE					
C	BUILDINGS & STRUCTURES					
D	PROCESS EQUIPMENT					
E	PIPING	10 047	197 125	164 640		361 765
F	ELECTRICAL					
G	PAINTING					
L	PLANT FACILITIES					
N	INSTRUMENTS & CONTROLS				60 320	60 320
P	INSULATION					
T	DEMOLITION					
	DIRECT FIELD COST	10 047	197 125	164 640	60 320	422 085
H	FIELD STAFF					
HB	LEGALITIES					
K	CONSTRUCTION SUPPLIES					
M	STARTUP					
S	TEMPORARY FACILITIES					
V	CRAFT BENEFITS					
V6	CONSTRUCTION CAMP.					
W	CONSTRUCTION EQUIP.					
	INDIRECT FIELD COST					181 515
	TOTAL FIELD COST					603 600
J	HOME OFFICE					42 205
	TOTAL FIELD & HOME OFFICE					645 845
Q	TAXES					11 045
R	PREMIUM PAY					3 940
Y3	CONTINGENCY					99 120
Y4	ESCALATION					
	SUB TOTAL					759 910
Y2	FEES					33 480
	TOTAL					793 390 *
	* $397.00 per Linear Foot					

THIS ESTIMATE IS BASED UPON CONTRACTOR'S EXPERIENCE, INCLUDING STATED RANGES OF ACCURACY. THE PROBABILITY OF PROJECT COST BEING WITHIN THE STATED ACCURACY RANGE IS ___%. CONVERSELY THE PROBABILITY OF PROJECT COST OUTSIDE THE STATED ACCURACY RANGES IS___%. THEREFORE, CONTRACTOR DOES NOT ASSUME THE RISK OF COSTS VARYING FROM THOSE ESTIMATED.

FORM 82-11FSC
(REPLACES FORM 82-11F)

Steams Catalytic

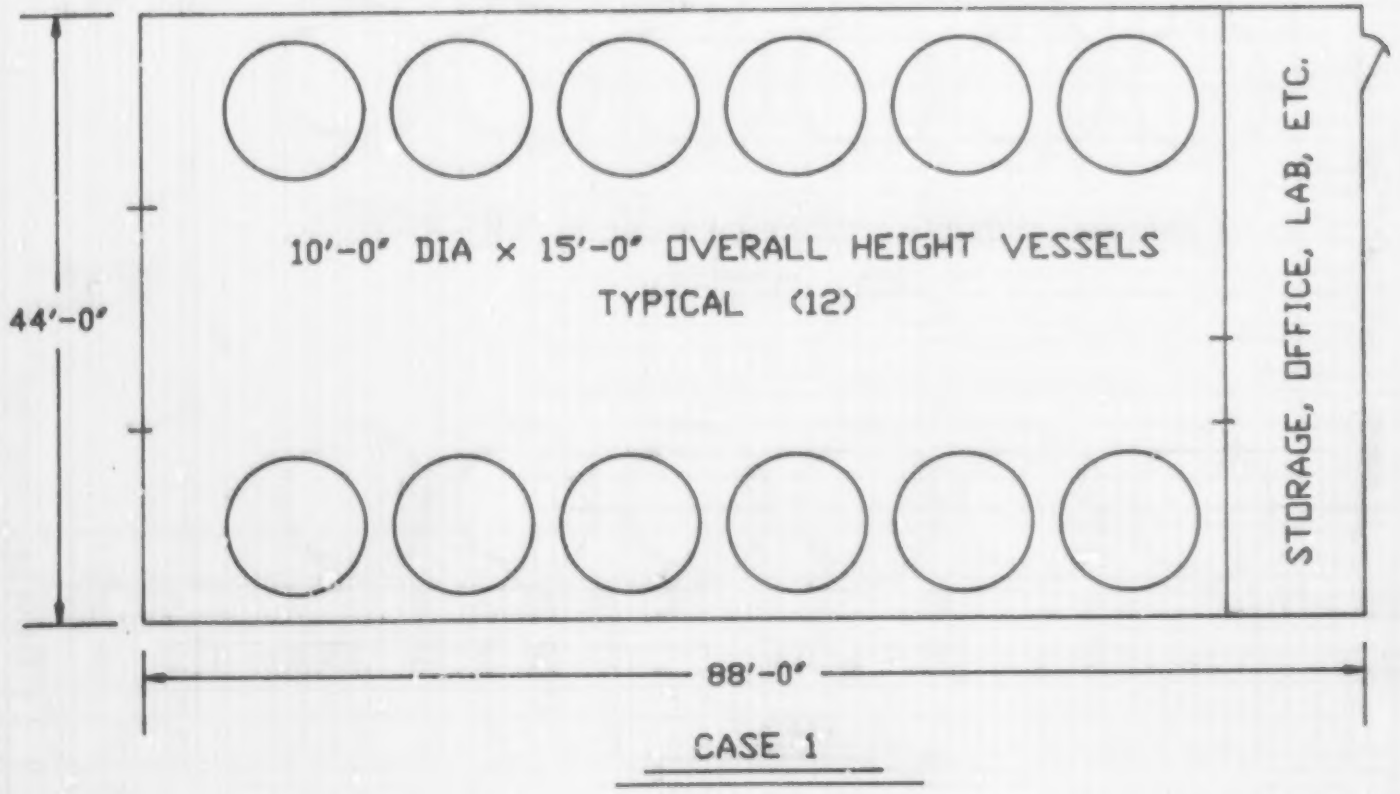
ESTIMATE SUMMARY

CLIENT	BuREC - Task 3	JOB NO	27835
LOCATION	Blythe, California	DATE	11-4-85
PROJECT	2000 Feet of 4" Pipe	BY	A. Cole
REV. NO.	REV. DATE	BY	

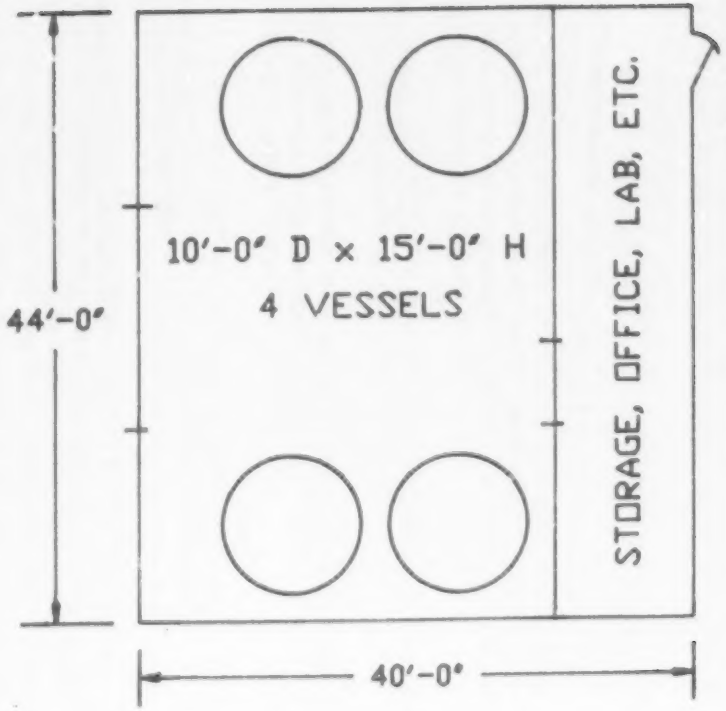
ACT	DESCRIPTION	CRAFT HOURS	LABOR	MATERIAL	OTHER	TOTAL	
A	EARTHWORK						
B	CONCRETE						
C	BUILDINGS & STRUCTURES						
D	PROCESS EQUIPMENT						
E	PIPING	1 141	22 390	21 790		44 180	
F	ELECTRICAL						
G	PAINTING						
L	PLANT FACILITIES						
N	INSTRUMENTS & CONTROLS						
P	INSULATION				16 690	16 690	
T	DEMOLITION						
	DIRECT FIELD COST	1 141	22 340	21 790	16 690	60 870	
H	FIELD STAFF						
HB	LEGALITIES						
K	CONSTRUCTION SUPPLIES						
M	STARTUP						
S	TEMPORARY FACILITIES						
V	CRAFT BENEFITS						
V6	CONSTRUCTION CAMP.						
W	CONSTRUCTION EQUIP.						
	INDIRECT FIELD COST					20 620	
	TOTAL FIELD COST					81 490	
J	HOME OFFICE					6 085	
	TOTAL FIELD & HOME OFFICE					87 575	
Q	TAXES					1 465	
R	PREMIUM PAY					450	
Y3	CONTINGENCY					13 240	
Y4	ESCALATION						
	SUB TOTAL					102 730	
Y2	FEES					4 525	
	TOTAL					107 255	*
	* $54.00 per Linear Foot						

THIS ESTIMATE IS BASED UPON CONTRACTOR'S EXPERIENCE, INCLUDING STATED RANGES OF ACCURACY. THE PROBABILITY OF PROJECT COST BEING WITHIN THE STATED ACCURACY RANGE IS ___%; CONVERSELY THE PROBABILITY OF PROJECT COST OUTSIDE THE STATED ACCURACY RANGES IS___%. THEREFORE, CONTRACTOR DOES NOT ASSUME THE RISK OF COSTS VARYING FROM THOSE ESTIMATED.

FORM 82-11FSC
(REPLACES FORM 82-11F)


10'-0" DIA x 15'-0" OVERALL HEIGHT VESSELS
TYPICAL (12)
STORAGE, OFFICE, LAB, ETC.
44'-0"
88'-0"

CASE 1



CASE 2

Section 9

GLOSSARY

Wastewater Evaporator - A vertical-tube, vapor compression falling film evaporator used for concentrating wastewaters and recovering a high quality product water.

Blowdown - An effluent stream from a cooling tower, softener, boiler or other such equipment or pond, which is used to control dissolved and/or suspended solids concentration in the source through partial removal of these solids.

Cycles of Concentration - The ratio of the concentration of conserved chemical species in the blowdown stream to the concentration of the same species in the feedwater.

Demineralizer - A water treatment system which uses ion exchange to removed dissolved salts from a water source.

Electrodialysis - A water treatment process which uses electrodes of opposite charge in conjunction with membranes which selectively remove positive and negative ions, resulting in a concentration of the feed stream and production of a relatively high quality product water.

Flue Gas Desulfurization (FGD) - A system used to reduce sulfur dioxide stack gas emissions from power plants using fossil fuels.

Ion Exchange Softening - A water treatment process which reduces calcium and magnesium by ion exchange with sodium and hydrogen ions, thereby reducing the scaling potential of a given feedwater source.

Levelized Cost - A uniform cost determined by reducing non-uniform costs to present worth by a specified discount rate, then multiplying the present worth by the capital recovery factor.

Levelized Capital Carrying Charge Rate - A fractional multiplier for the capital investment which results in the levelized annual revenue requirement necessary to meet the fixed costs associated with the investment. The fixed costs typically include the cost of capital, depreciation, property taxes, insurance, allocation of income taxes and associated administrative burden.

Lime/Soda Softening - A process which reduces calcium and magnesium through precipitation reaction with lime and soda ash.

Reactor/Clarifier (Reactivator) - A vessel which provides retention time for the completion of lime or lime/soda softening reactions, settling of the precipitated solids and clarification of the effluent stream.

Reverse Osmosis (RO) - A water treatment process which uses semi-permeable membranes operated under hydraulic pressure to concentrate feedwater and produce a relatively high quality product water.

Stoichiometric - Chemical substances which are in exact proportions for a given reaction.

Surge Pond - A pond which holds waters prior to their transfer to a reuse or processing system.

Section 10

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